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INVESTIGATION OF THE USE OF HIGH PURITY OXYGEN AERATION IN THE CONVENTIONAL ACTIVATED SLUDGE PROCESS

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George Knapp 9-21-70

Institution
U.S. Geological Survey, Water Resources Div.

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INVESTIGATION OF THE USE OF HIGH PURITY OXYGEN AERATION IN THE CONVENTIONAL ACTIVATED SLUDGE PROCESS

by

Mr. J. G. Albertsson
Dr. J. R. McWhirter
Dr. E. K. Robinson
Mr. N. P. Vahldieck

Union Carbide Corporation
Linde Division
Tonawanda, New York

for the
FEDERAL WATER QUALITY ADMINISTRATION
DEPARTMENT OF THE INTERIOR

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FWQA Project Officer, R. C. Brenner
Advanced Waste Treatment Research Laboratory
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ABSTRACT

A full scale system designed to demonstrate the practical and economical use of high purity oxygen gas in aeration of the conventional activated sludge waste treatment process has been tested under a variety of operating conditions in a direct performance comparison with a parallel air aerated system. A sparged-turbine, gas-liquid contacting unit was employed in covered tanks of conventional design with discreet liquid and gas staging for cocurrent flow of each phase. This oxygenation system required very low power input for oxygen transfer and liquid mixing (0.08 - 0.140 HP/1000 gal. mixed-liquor) to routinely operate at dissolved oxygen concentrations of 8-10 mg/l while achieving > 90% utilization of feed oxygen gas. It was shown that the oxygenation system could operate with MLVSS concentrations of as high as 4500 mg/l achieving about 90% BOD removals at aeration detention times as low as 1.2 hours (raw flow + recycle flow) treating domestic waste of 220 mg/l BOD. Under other conditions and in direct comparison with air aeration, the oxygenation system consistently exhibited treatment performance superior to that achieved with air aeration even at uneconomically high aeration rates (3-5 cf air/gal. waste treated). Biomass from the oxygenation system is highly flocculent and readily settleable with desirable handling characteristics. Recycle and waste activated sludge suspended solids concentrations of about 3% were achieved.

Process advantages recognized with the oxygenation system indicate significant reductions in secondary waste treatment costs in comparison to conventional diffused air aerated systems. These advantages are principally reduced aeration tank volume requirements for equivalent treatment and a reduced production of waste activated sludge, a result of more efficient utilization of high purity oxygen than here-to-fore possible. As an example, the total cost of treatment, including preliminary treatment, with air at 6 hours nominal detention time and a 1.6 cf air/gal. aeration rate is estimated to be 15.1 cents, 13.0 cents and 11.0 cents/1000 gal. for plant sizes of 6, 30, and 100 MGD respectively while these costs are estimated to be reduced to 11.9 cents, 9.5 cents, and 7.8 cents/1000 gal. for respective 6, 30 and 100 MGD plant sizes with the use of an oxygenation treatment system capable of equivalent BOD removal at 2 hours nominal treatment detention time.

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CONCLUSIONS AND RECOMMENDATIONS

A practical and economically attractive means for use of high purity oxygen gas in aeration of the conventional activated sludge process has been tested over an eight month period of continuous operation. This work was done using a temporarily installed oxygenation system in the existing aeration tanks at the Batavia, New York Municipal Pollution Control Plant. A sparged-turbine, gas-liquid contacting unit was employed for oxygen transfer. The aeration tanks,which were of conventional design,were covered and equipped with gas-liquid staging baffles to provide a multistage oxygenation system. This system functioned with co-current gas and liquid flow.

The test program was specifically planned to evaluate the following:

1. The feasibility and economics of high purity oxygen use in aeration of the conventional activated sludge waste treatment process.
2. The comparison of oxygenation and air aeration in terms of treatment performance.
3. The consistent operation of the oxygenation system at high mixed-liquor suspended solids (MLSS) levels (6000-8000 mg/l), high dissolved oxygen concentrations (8-10 mg/l) and high overall utilization of feed oxygen gas (>90%).
4. The operation and treatment performance of an oxygenation system at low detention times (approximately 1.2 hours based on raw wastewater + recycle sludge flow) and high organic loading conditions considered economically desirable but impractical with air aeration.
5. The relative economics of domestic waste treatment comparing the cost of oxygenation with conventional (diffused air) air aeration.

The work was conducted in three phases. In two phases,the performance of the oxygenation system was compared to a parallel air aeration system. In a third phase, the oxygenation system was used with one-fourth of the plant aeration tankage to treat the entire wastewater flow to the plant. A summary of results from each of the three phases of operation is shown in Table 1.

TABLE 1

SUMMARY COMPARISON OF AIR AND OXYGENATION SYSTEMS PERFORMANCE

	PHASE I OPERATION		PHASE II OPERATION	PHASE III OPERATION	
	AIR	OXYGEN	OXYGEN	AIR	OXYGEN
Wastewater Feed Rate (MGD)	1.97	1.91	2.53	1.29	1.44
Aeration Detention Time (Hrs.)*	3.6	3.6	1.2	2.6	2.0
Nominal Aeration Detention Time (Hrs.)**	4.0	4.1	1.5	3.0	2.8
MLSS Concentration (mg/1)	2440	3060	6980	3640	6190
MLVSS Concentration (mg/1)	1740	2210	4450	2580	4310
Recycle Sludge TSS (mg/1)	14,960	18,620	29,560	16,600	18,790
Volumetric Organic Loading lbs. BOD/Day/1000 ft^3	60.0	57.9	212.5	128.9	144.8
Mixed-Liquor Dissolved Oxygen Concentration (mg/1)	1.5	8.7	9.0	0.8	8.0
Ft^3 Air Utilized/Gal Sewage Treated	2.89	-	-	4.32	-
% Feed Oxygen Utilized		95.5	92.7	-	91.4
% BOD Removed	90	92	90	88	94
% COD Removed	76	80	71	79	84
% TSS Removed	93	96	89	94	97

* Raw Flow + Recycle Flow

** Raw Flow Only

The oxygenation system consistently exhibited superior treatment performance in direct comparison to the parallel air aeration system in spite of the fact that an impractically high air aeration rate was used to insure optimum performance. The power required by the oxygenation unit to accomplish high treatment efficiency was only 0.08 - 0.14 HP/1000 gal. of wastewater treated.

The biomass generated by the oxygenation system was highly flocculant and rapidly settleable, yielding sludge volume index values as low as an average of 36. The high recycle and waste solids total suspended solids concentration achieved (approximately 3-4% when clarifier operation was closely controlled) obviates the requirement for thickening of the waste activated sludge prior to further processing for disposal.

An extensive evaluation of the oxygenation system treatment costs in comparison to conventional air aeration reveals significant potential savings attributable to high purity oxygen use. These savings are due to several characteristics of the oxygenation system, but the major factors are reduced aeration tank volume requirements for equivalent treatment and a reduced production of waste activated sludge. As an example, the total cost of air aeration treatment at 6 hours nominal detention time (based on raw wastewater flow) and a 1.6 cf/gal aeration rate is estimated to be 15.1 cents, 13.0 cents, and 11.0 cents/1000 gal for plant sizes treating 6, 30, and 100 MGD of wastewater. For equivalent BOD removal with oxygen aeration at 2 hour nominal detention time (based on raw wastewater flow) these total treatment costs are estimated to be 11.9 cents, 9.5 cents, and 7.8 cents/1000 gal for the respective 6, 30, and 100 MGD plant sizes. Total treatment costs referred to here include investment and operating costs associated with preliminary and primary treatment as well as the investment and operating costs associated with secondary treatment by the activated sludge process and the disposal of primary and waste activated sludges.

The promising results of the work reported clearly indicate the advisability of further work to evaluate in detail certain characteristics of the oxygenation system not yet studied in sufficient depth. Specifically, the existing temporary installation should be used to evaluate the following:

1. The biokinetics of the treatment process as revealed by careful control of sludge wasting, frequent analysis of soluble BOD and COD in feed wastewater, mixed-liquor filtrate and final clarifier effluent.

2. Flow control should be established for the system such that the measurements in (1) above can be made under more nearly steady state conditions.

3. The handling properties of the waste activated sludge should be more fully elucidated by direct vacuum filtration tests.

4. System performance should be optimized by installation and operation of a flow proportion controlled activated sludge return system.

5. It would be desirable to evaluate oxygen aerated, aerobic digestion of waste activated sludge and to establish the handling characteristics of such stablized sludge by vacuum filtration tests.

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J. R. McWhirter - Product Manager, New Products
F. W. Bonnet - Technology Manager
E. K. Robinson - Contract Manager
J. G. Albertsson - Supervisor
N. P. Vahldieck - Senior Engineer
R. J. Grader - Staff Engineer
D. V. Daly - Technician
J. R. Duemmer - Technician
S. E. Faruga - Technician
E. H. Kremer - Technician
P. F. Layer - Technician

Appreciation is expressed to the City of Batavia, New York for permission to conduct the work described here at their Municipal Pollution Control Plant. The enthusiastic attitude displayed toward this experimental program by Ira M. Gates, City Manager; Robert Lawrence, Superintendent of Water and Sewage Works and his competent staff contributed greatly to the successful completion of the program.

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BACKGROUND AND INTRODUCTION

Historical Aspects

The use of pure oxygen or oxygen enriched air in the activated sludge waste treatment process has been the subject of a number of technical and economic studies published over a period of more than twenty years. Okun(1) first reported on laboratory tests using oxygen in a modification of the conventional activated sludge process in 1949. This process, termed the "bio-precipitation system", was developed from a method first suggested by Pirnie(2). Okun reported the bio-precipitation process to be an effective means of treating wastewater, at least equally efficient as present day completely mixed activated sludge systems. In this process mixed-liquor is not directly aerated, but rather the oxygen required is predissolved in the feed wastewater to the bio-oxidation (mixed-liquor) reactor. The bio-oxidation takes place in a gently agitated fluidized bed system which acts as an integral reactor and clarifier. To minimize the costly recycle of wastewater through the sludge bed to supply required oxygen, the dissolved oxygen content of the feed stream to the reactor was maximized using pure oxygen.

Unfortunately, the dissolved oxygen content of this stream must be raised to a very high value and thus the effective mass transfer driving force for oxygen dissolution into the liquid is not significantly greater than that for conventional air aeration practice, even though a high oxygen partial pressure aerating gas is used. This process arrangement also is not consistent with achievement of an economically desirable high percentage oxygen absorption and utilization necessary to minimize oxygen production costs. Consequently, the bio-precipitation process has certain inherently undesirable features which tend to increase the cost of supplying oxygen to the reactor relative to the cost of direct aeration or oxygenation of the mixed-liquor itself.

Bud and Lambeth(3) studied the bio-precipitation process in a large pilot plant. Using pure oxygen, they reported an oxygen absorption efficiency of 20 to 25% with an exhaust gas composition of 45-50% oxygen. The reported power requirement, including oxygen generation, was equal to conventional air aerated activated sludge processes. A potential reduction in size of aeration tankage and final settler volume was indicated as a distinct advantage for the process in terms of investment and maintenance costs. These potential savings were based, however, on a conservatively designed conventional activated sludge system employing six hours aeration detention time (based on raw waste flow).

Okun and Lynn(4) in later studies reported the sole advantage of oxygen aeration was to reduce or eliminate the period of zero D.O. (dissolved oxygen) yielding an apparent, but not real, increase in effective sludge activity. They concluded that substitution of oxygen for air or use of oxygen enriched air in standard activated sludge aeration tanks would not be economically feasible because of poor oxygen transfer efficiency inherent in diffuser and tank design.

Both Okun(5) and McKinney(6) have reviewed the "state-of-the-art" in the use of high purity oxygen in secondary treatment and have recognized the inherent advantage of an aerating gas with a high oxygen partial pressure in increasing the rate of oxygen transfer. Carver(7) has reported data indicating that the rate of oxygen transfer with pure oxygen aeration was essentially independent of dissolved oxygen content of the liquid between 0 and 12 mg/l. He suggested that for systems exerting a high oxygen demand, additional oxygen might be furnished as pure oxygen and some air aeration might be used for providing adequate mixing only.

McKinney(6) concludes that several advantages may be achieved if pure oxygen or oxygen enriched air could be used in the activated sludge process. These would be the following:

1. Possibility of avoiding excessively high aeration rates and hence obtaining a reduction in the power required per unit of oxygen transferred.
2. Increased rate of stabilization of organic material.
3. Reduction in, or elimination of periods of zero dissolved oxygen concentration.
4. Ability to operate high rate systems by substantial increases in organic loading where oxygen is not limiting.
5. Reduction in plant size and thus capital investment.
6. Increased capacity of organically overloaded plants without need for additional aerator capacity.

To this time, the major problem in realizing these potential advantages has been the development of an efficient means of utilizing oxygen such that a high overall utilization of gaseous oxygen, as great as 90%, might be achieved. It has been apparent that development of satisfactory means for effectively and efficiently utilizing high purity oxygen in the activated sludge process was the lacking process ingredient. The absence of this element of the technology appears as the main factor in generating the current widespread impression that oxygen use must, of necessity, be significantly more costly than air aeration.

Air Aeration Technology

Devices currently used in the air aeration of mixed-liquor bio-oxidation processes can be broadly categorized in three main classes. These are diffused air systems, submerged turbine units, and surface aeration units. These three aeration methods span the complete spectrum of means for supplying the necessary energy for mass transfer of oxygen from the gas to the liquid phase and for mixing and contacting of the wastewater with the activated sludge biomass. On one end of the spectrum diffused air systems supply all input energy by gas compression to generate the interfacial area and turbulence for interphase oxygen transfer and to accomplish adequate fluid motion in the mixed-liquor for homogenous suspension of solids. At the opposite end of the spectrum surface aerators supply all energy input directly to the liquid phase by mechanically spraying large quantities of mixed-liquor into the open atmosphere to promote oxygen transfer in addition to bulk liquid mixing. Submerged turbine units combine the features of the previous devices by utilizing energy for both gas compression and mechanical fluid mixing. Although each of these three types of aeration devices with some variations are in general use, diffused air systems seem to predominate, especially in municipal treatment plants using the activated sludge process.

In spite of the apparently basic differences among the three types of air aeration devices described, the available literature indicates that their effective oxygen transfer efficiency (as lbs oxygen transferred/ HP hour) differs by only about 25 to 30% in normal practice. This, no doubt, takes into account the variations introduced from procedures by which the performance of air aeration devices are traditionally evaluated experimentally(8,9). A host of confusing claims and apparent anomalies are reported in the literature which often are merely the result of differing testing techniques, aeration conditions, and the particular basis on which the performance is calculated and reported. Considerable care must, therefore, be exercised in evaluating energy requirements and costs associated with supplying oxygen to activated sludge processes using standard air aeration techniques. These factors, however, have little influence on the relative performance comparison of air and oxygen aeration systems.

One of the generally unrecognized aspects of air aeration is the relatively high energy requirement for dissolution of oxygen into the aqueous mixed-liquor suspension. This is a consequence of the basic mass transfer process which is in this instance almost entirely liquid phase controlled, a distinguishing feature of the dissolution of all sparingly soluble gases in liquids. As a result, large quantities of energy are expended in creating liquid surface and interfacial turbulence to enhance interphase mass transfer rates. In spite of considerable effort and technological advances made over the years, however, the gas-liquid contacting and oxygen dissolution step remains an inherently inefficient process in terms of the quantity of oxygen dissolved per unit of energy input. This point is best illustrated in Table 2 wherein the energy requirement of oxygen mass transfer in diffused air aeration is shown as a function of the aeration conditions and mixed-liquor dissolved oxygen level. These energy requirements, expressed as KWH/100 cf of oxygen dissolved,are substantially greater than the cost of high purity oxygen production (1.4 KWH/100 cf) by today's standard air separation processing techniques. The power required, therefore, to separate oxygen from air is considerably less than that required to dissolve oxygen from air into mixed-liquor systems. This is the case even at zero dissolved oxygen concentrations, an unacceptable low value for the activated sludge process by today's standards. Therefore if the energy expended to dissolve high purity oxygen in mixed-liquor is equal to or less than the difference between the energy required to transfer oxygen directly from air and the energy required to produce high purity oxygen by air separation, then pure oxygen use may be economically competitive with air for secondary treatment aeration.

Although standard design practice calls for mixed-liquor dissolved oxygen concentrations of 1 to 2 mg/l, several reports in the literature have indicated significant process performance and treatment quality advantages in activated sludge treatment at dissolved oxygen levels of 4-5 mg/l or in some instances greater. As Table 2 clearly illustrates, the unfortunate situation is that the energy required, hence the cost, of oxygen dissolution with air aeration,increases sharply with increase in the sustained bulk liquid dissolved oxygen level. This is not unexpected since at atmospheric pressure and 20°C the oxygen equilibrium concentration of water with air is approximately 9.0 mg/l, equivalent to a partial

TABLE 2

DIFFUSED-AIR AERATION ENERGY REQUIREMENTS AS A FUNCTION OF AERATION CONDITIONS*

Liquid Temp. (°C)	Mixed-Liquor D.O. (mg/1)	Effective Transfer Eff. (Lbs. O_2/HP Hr.)	Energy Consumption (KWH/100 CF)
20.0	0.0	1.67	3.71
20.0	1.0	1.48	4.19
20.0	2.0	1.29	4.80
20.0	3.0	1.10	5.73
20.0	4.0	.91	6.81
20.0	5.0	.72	8.61
20.0	6.0	.53	11.7
30.0	0.0	1.76	3.52
30.0	1.0	1.52	4.07
30.0	2.0	1.28	4.85
30.0	3.0	1.04	5.96
30.0	4.0	0.795	7.80
30.0	5.0	0.557	11.1
30.0	6.0	0.317	19.5

* Aeration conditions and equipment efficiency assumptions

Alpha factor - 0.80
Beta factor - 0.95
Blower motor efficiency - 0.90
Compressor efficiency - 0.70
Compressor discharge pressure - 8.1 psig

pressure of oxygen in air of 160 mm Hg. Consequently, the energy required (and capital investment in equipment) increases by a large factor to increase mixed-liquor residual dissolved oxygen levels from 1.0 to 5.0 mg/l. To minimize cost of plant construction and operation, design of aeration systems often is planned for very low dissolved oxygen levels. It is clear, however, that treatment processes designed to operate on the threshold of oxygen limiting conditions for the biomass may significantly reduce the ultimate effectiveness of activated sludge treatment. In the following, the basis for a practical and economical method of activated sludge aeration, the use of high purity oxygen, will be contrasted to illustrate how operation at dissolved oxygen levels of 5-10 mg/l may be accomplished at costs competitive with conventional air aeration at dissolved oxygen levels of 1-2 mg/l.

Oxygen Aeration System Requirements

As pointed out in the foregoing, the potential for the use of high purity gaseous oxygen in biological waste treatment processes, and by the activated sludge process in particular, is a relatively well known idea. The manner in which oxygen can be effectively and economically employed in such processes is not, however, obvious. This statement is supported by the fact that, in spite of more than two decades of interest in oxygen use, no one has yet proposed (or at least reported) an effective and economical means for doing so. Under the illustrative conditions of Table 2, the power requirement to maintain 1.0 to 5.0 mg/l D.O. at 20°C with air aeration is 3.0 to 6.2 times greater than the power required for high purity oxygen production by air separation. Therefore, oxygen use may be economically competitive with air aeration if a means is available so that oxygen in a relatively pure state can be efficiently (in terms of power requirement) and effectively (in terms of high percentage absorption) dissovled in mixed-liquor.

As a first approximation, the efficiency of the dissolution process should be higher with oxygen as the aerating gas merely as a result of the increase in the partial pressure driving force for interphase transfer (approximately 760 mm Hg for pure oxygen and approximately 160 mm Hg for air - a 4.75 fold increase). At the same time it is clear that high energy utilization efficiency is not the only consideration. Simultaneously, a high degree of oxygen dissolution and utilization is necessary. Consequently, the problem of oxygen use is one of achieving a high percentage absorption while maintaining a high energy utilization efficiency for oxygen dissolution and for mixing of the activated sludge biomass suspension.

It is well known that air aeration systems achieve only a relatively small percentage absorption of the oxygen bubbled through or otherwise exposed to the mixed-liquor. Diffused air systems, for example, achieve only about a 5% absorption of the oxygen content of air while submerged turbine units operate at 15-20% oxygen absorption efficiency. Since, as stated previously, an oxygen aeration system must be extremely efficient in terms of overall oxygen absorption and utilization to be competitive with air, an analysis of conventional air aeration techniques reveals that these approaches are not well suited to oxygen use.

Consideration of the basic requirements of an oxygen aeration system reveals that the fundamental gas-liquid contacting problem and basic mass transfer process must, indeed, be quite different from those of air systems. Table 3 lists the basic requirements to be met in oxygen aeration system design and the unique characteristics relative to air aeration. The more important factors are high percentage oxygen absorption and energy utilization efficiency, the small volume of gas to be contacted and the multi-component character of the basic mass transfer process itself. These requirements support the contention that aeration methods effective (optimal) for air systems will not be effective or optimal for an oxygen aeration system and vice versa.

Linde Oxygen System Technology

Recognizing that cost effective activated sludge oxygen aeration systems must by definition differ from conventional air systems in the means used to transfer oxygen, research and development activities have been conducted over a six year period at the Union Carbide Corporation's

TABLE 3

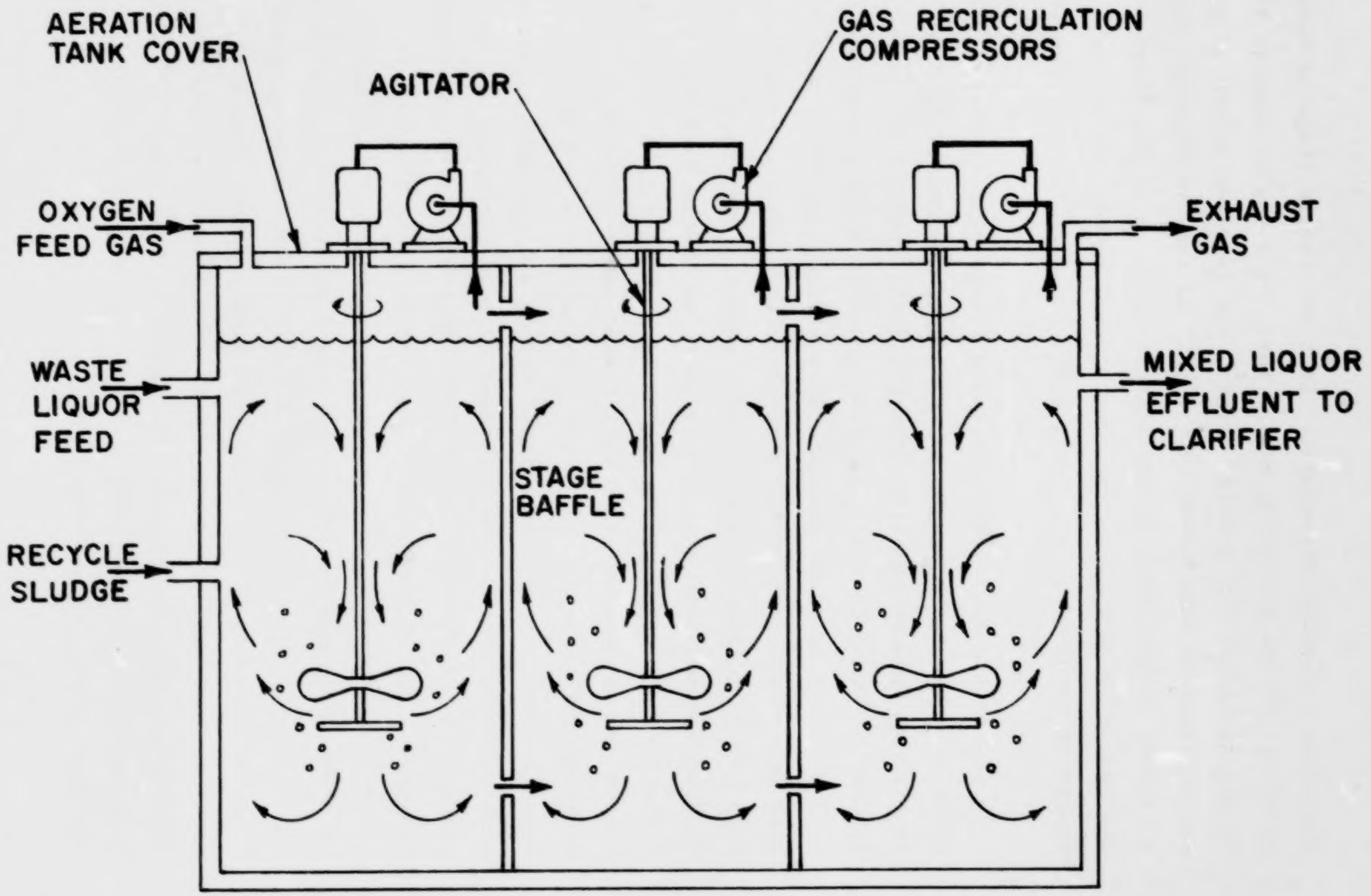
BASIC REQUIREMENTS AND CHARACTERISTICS OF OXYGEN AERATION SYSTEMS

1. Enable high percentage absorption and utilization (>90%).
2. Obtain high overall energy transfer efficiency in the dissolution process as compared with air aeration systems.
3. Provide ample liquid mixing.
    - A. Maintain sludge suspension
    - B. Maintain uniform mixed-liquor composition
4. Simple design with low maintenance costs
5. Low capital investment.
6. Volume of gas to be contacted with the liquid is about 1/90 that of typical diffused air aeration systems.
7. Associated with (6) above, mechanical energy input will be required to achieve the required level of bulk fluid mixing and circulation.
8. Multi-component mass transfer process. Mass transfer of inert gases (such as nitrogen, argon, and water vapor) and bio-oxidation reaction products (carbon dioxide) must be taken into account.

Linde Division to devise simple yet highly efficient oxygenation systems and processes applicable and competitive in the waste treatment field. The result of this work was the development of several oxygen aeration techniques, systems and process concepts applicable to the activated sludge treatment of wastewater. One such method and systems approach, having particular merit, was demonstrated in a full scale application in the FWQA funded contract work reported here. Patents for the oxygenation system technology described in this report have been applied for by the Linde Division of Union Carbide Corporation. This particular system is based upon the use of a series of cocurrent gas-liquid contacting stages to enable high overall oxygen absorption efficiency at a high average energy transfer efficiency. Figure 1 shows a schematic diagram of the system. A series of sparger-turbine gas-liquid contacting units are employed. This conceptual drawing illustrates four such stages in series, but the number of stages may vary from one application to another. Details of the gas-liquid contacting unit design employed in the work described here will be given in the section to follow. In such a system as shown in Figure 1, the oxygen gas is fed into the first stage at a pressure of only about 1-3 inches of water above ambient. Small recirculating gas blowers in each stage pump the oxygen through a hollow shaft to the rotating sparge device at a rate sufficient to maintain mixed-liquor D.O. through oxygen mass transfer. The indicated pumping action of the impeller located on the same shaft as the sparger promotes adequate liquid mixing and yields relatively long residence times for the dispersed oxygen bubbles. Gas is recirculated within a stage at a higher rate than the rate of gas flow from one stage to another. The successive aeration stages or chambers are connected to each other in a manner which will allow gas to flow freely from stage to stage with only a minor pressure drop (<1.0 inches H_2O) but yet sufficient to prevent gas back mixing or interstage mixing of the aeration gases. This is accomplished by appropriate sizing of the interstage gas passages. The liquid flow (mixed-liquor) through successive stages is co-current with gas flow. Stage liquid volumes are sized appropriately to yield desired aeration detention times

FIGURE I

SCHEMATIC DIAGRAM OF MULTI-STAGE OXYGENATION SYSTEM


AERATION TANK COVER
AGITATOR
GAS RECIRCULATION COMPRESSORS
OXYGEN FEED GAS
EXHAUST GAS
WASTE LIQUOR FEED
MIXED LIQUOR EFFLUENT TO CLARIFIER
STAGE BAFFLE
RECYCLE SLUDGE

dependent on design throughput. Each successive stage is essentially identical to the preceding except that, as a higher portion of the oxygen demand is met in the initial stages, the required volume of gas to be recirculated in subsequent stages will be less to maintain the desired dissolved oxygen level in the mixed-liquor. Effluent mixed-liquor from the system is settled in the conventional manner and the settled activated sludge is returned to the first stage for blending with feed raw sewage.

The entire multistage activated sludge unit is fitted with a gas tight cover to contain the oxygen aeration gas, the only exception being a restricted exhaust gas line from the final stage. Gas is fed to the system on demand (a small positive pressure being maintained by a flow controller to overcome pressure drop through the unit) with the entire unit operating in effect as a respirometer. As the organic load and respiration (oxygen demand) of the biomass increases,the pressure tends to decrease and feed oxygen flow into the system increases to re-establish a pressure set point of the controller. Feed oxygen to the multistage system can be controlled on this demand basis by a simple regulator or differential pressure controller - automatic valve combination. The overall power required for liquid mixing and gas recirculation will vary with specific system configurations, but in virtually all cases will lie between the limits of 0.08 and 0.16 HP/1000 gal. of mixed-liquor under aeration. On the same energy requirement basis as shown for air aeration in Table 2, this power input corresponds to about 1.2 KWH/100 cf of oxygen transferred. The total power required to generate and dissolve 100 cf of oxygen would be about 2.6 KWH in most cases (1.4 KWH is required to generate 100 cf of high purity oxygen from air). As may be noted by comparison of these oxygen system power requirements with those for air aeration in Table 2, substantially less power is required to separate oxygen from air and dissolve it in the mixed-liquor than to dissolve oxygen from air directly.

In general the net flow rate of gas from stage to stage is largely determined by the net rate of oxygen mass transfer to the liquid in each stage. Since the rate of oxygen transfer to the liquid is usually higher than the desorption rate of nitrogen and carbon dioxide from the mixed-liquor, the gas flow rate will usually decrease from stage to stage. Concurrently, however, the oxygen partial pressure of the gas phase in successive stages will gradually decrease as the CO_2 and nitrogen content increases. Normally the system will operate best with a gas composition in the final stage of about 50% oxygen, corresponding to a 90-95% overall oxygen absorption efficiency. Such efficiency is well within the range of economic usage. The overall percentage oxygen absorption and average oxygen transfer efficiency for the multistage system is determined by the feed gas rate, number of stages, and gas-liquid contacting effectiveness of the aerator or aerator energy input per stage.

It is necessary to vent gas from the final stage or the inert gas concentration buildup throughout the successive stages would rapidly reach levels unacceptable for efficient oxygen transfer. The rate of venting will, in addition, depend upon the rate of oxygen feed, thus upon the set point of the feed oxygen controller. Due to a net dissolution of oxygen, the vent gas rate will be only a small fraction, e.g. about 10-20% of the oxygen feed rate.

In spite of the dependency of overall efficiency of oxygen transfer on the number of successive system stages, it is a consequence of system parameters that a relatively small number of stages can achieve a high level of performance. As an example of this consideration, one may compare the average energy transfer efficiency of multistage systems for equal total horsepower at a constant feed oxygen rate. The result of this comparison indicates that a single stage system would require twice the power of a six stage unit to achieve a 90% oxygen utilization efficiency. This is one of the most significant economic implications of the described multistage aeration system; high overall oxygen absorption efficiencies at economically attractive power input levels.

One underlying reason for the increased performance capability of multistage systems is, that for a given level of oxygen absorption or total power input, the oxygen transfer is occurring at an average partial pressure driving force which is higher than that obtainable in a single stage unit. A single stage contacting chamber wherein the gas phase is completely mixed cannot operate at both a commensurately high percentage absorption and a high oxygen partial pressure. One must necessarily be obtained at the expense of the other. The same general compromise, of course, also applies to multistage systems, but considerably higher overall average performance levels can be achieved.

A very desirable feature of a multistage contacting system is that it lends itself very well to simultaneous staging of the mixed-liquor as well. The potential benefit of a liquid staged reaction system to improve treatment quality per unit volume of mixed-liquor under aeration has been reported by many investigators. Pipes et al (10) have evaluated staging effects both theoretically and experimentally. Only a relatively small number of stages gives reasonably good results in relation to an ideal infinite stage or plug flow system.

It is well known that in a plug flow or multiple liquid stage activated sludge system, the oxygen demand varies considerably from the feed end to the effluent end of the system. Staging increases the oxygen demand at the liquid feed end of the unit, compared to a completely mixed system. Others have found that an oxygen deficit exists at the influent end of the plug flow tanks for diffused air systems. The oxygen transfer capacity of a multistage cocurrent oxygen aeration system, however, naturally varies from stage to stage with the transfer rate decreasing from the feed stage to the final or exhaust gas stage, as a result of decreased gas phase oxygen composition. Thus, the use of cocurrent gas and wastewater flow through a multistage contacting system tends to match the natural oxygen demand requirement variation of the mixed-liquor with the oxygen transfer capactiy variation of the gas contacting system. Of course, the individual transfer capacity per stage can also be further varied in design by changing the energy input per stage and the oxygen gas feed rate to the entire system.

Another very important aspect of oxygen aeration compared to air aeration is that the oxygen transfer capacity is considerably higher per unit of energy input. As discussed previously, this is a necessary requirement from economic considerations, but it also has important process performance implications. This means that the bulk liquid agitation and mixing rate can be controlled at lower levels while simultaneously obtaining higher D.O. concentrations. This is believed to be a very important characteristic required for the growth of a highly flocculant, readily settleable sludge mass particularly at high solids loadings. Thus, with oxygen aeration, the agitation rate can be controlled over wide ranges and at lower bulk liquid turbulence and shear levels to promote the growth of large floc particles and a readily settleable biomass. The micro-organisms will still function very efficiently in this state, however, because of the high bulk liquid D.O. levels (5-10 mg/l) which are economically attainable with oxygen aeration.

The use of multistage oxygen aeration also lends itself quite readily to conversion of existing air aeration systems (diffused-air rectangular aeration tanks). This has particularly important economic implications for the conversion of existing overloaded air aeration systems to oxygen use, and thus enable both increased capacity and treatment quality improvement.

Operation at High D.O. Levels

As stated previously, one of the most important advantages of oxygen aeration is the capability for economical operation at high D.O. levels compared to those which can practically be obtained for comparable or considerably higher cost with air aeration. An increase in the D. O. level from 2.0 to 8.0 mg/l results in a very small decrease (< 5%) in the energy transfer efficiency and in the overall percentage oxygen absorption for reasonable operating conditions. The performance data to be presented here shows that economical performance levels can be obtained at D.O. levels at least as high as 10.0 mg/l.

The cost penalty associated with a change in oxygenation system D.O. levels from 2.0 to 8.0 mg/l is quite small representing only a small percentage change in the mass transfer driving force. Quite the opposite is, of course, true for air aeration systems wherein a change in D.O. level from 2.0 to 5.0 mg/l represents about a 50 percent reduction in the available driving force for mass transfer. Table 2 indicates that this change in D.O. level would more than double the dissolution energy requirements for diffused-air systems as well as substantially increase the required capital investment. Information presented here will show that such is not the case for multistage oxygen aeration systems. Thus, it will be seen that with oxygen aeration, practical operation of the activated sludge process at high D.O. levels can be realized at costs which are economically competitive with air aeration at low D.O. levels. As stated earlier, this study has concluded that operation under these conditions can result in substantial treatment quality improvements as well as significant plant size reductions by enabling operation of high solids, high rate activated sludge systems.

The conclusions from the foregoing is that a multistage oxygen aeration activated sludge treatment system might be operated at controlled D.O. of 8 mg/l, for example, by varying feed oxygen input or the energy input into oxygen mass transfer regardless of organic loading variation where appropriately designed aerators are installed to meet peak oxygen demand of a given treatment system. Simple feed back control systems are possible where power input (volume of gas recirculated) might be varied to meet organic load. In some domestic treatment facilities,low power inputs may suffice at night with higher values for day loadings such that power conservation is possible while D.O. remains constant throughout the mixed-liquor at all times. It is important to note, however, that a major advantage of the oxygenation system described here is the ability to operate at high D.O. concentrations using only gas space pressure control as the basis for meeting oxygen demand rather than less reliable D.O. level control. The pressure control system is highly reliable and provides nearly instantaneous response to changes in oxygen demand as may result from hydraulic or organic load variations. A D.O. level control system may be used as a secondary means of control, but should be regarded as a non-essential option in most cases.

The described ability to operate at high D.O. without oxygen limitation permits a multistage oxygen aeration system to operate at high MLVSS levels (4000-5000 mg/l routinely). From a system kinetics point of view, one benefit is that high overall organic removal efficiency is possible at very low treatment detention times not possible with air. It is not practically feasible in conventional (normal organic loading) practice to operate an air aerated unit at MLVSS concentrations in excess of 2500-3000 mg/l at D.O. levels where oxygen limitation is not acute (approximately 1.0 mg/l). Impractically high aeration rates and power consumption would be necessary. A highly flocculant biomass enables the achievement of such MLVSS levels in the oxygenation systems at low sludge recycle ratios.

With the foregoing considerations in mind, the multistage oxygen aeration system described herein was developed and tested at the pilot scale. Results of these tests supported the contentions set forth here. As a result the Federal Water Quality Administration of the United States Department of the Interior funded an experimental program designed to test the oxygen system in a full scale domestic waste treatment plant under a variety of operating conditions. In this manner, data were obtained by direct comparison with the performance of a parallel conventional air aeration system to examine feasibility of oxygen use, overall treatment performance, and the overall economic value of the described system. The results of this work and conclusions evolved from these results are the subject of this final report for the project.

CONTRACT PROGRAM PLAN

The experimental program described in the following was undertaken to directly compare the relative process and economic performance of air and direct oxygen aeration in the conventional activated sludge process treating typical municipal sewage under practical operating conditions. The project was carried out at an existing municipal treatment plant enabling direct side by side comparison of the two aeration processes operating on the same feed wastewater under controlled aeration conditions in each case. The specific program objectives were the following:

1. Demonstration of a practical and economically attractive high purity oxygen gas-liquid contacting system for oxygenation of activated sludge mixed-liquor in biological treatment of raw wastewater.
2. Comparison of process treatment effectiveness and economics of an air aerated system operating at nominal dissolved oxygen levels (1.0-2.0 mg/l) with an oxygenation system operating at high dissolved oxygen concentrations (7-10 mg/l) under approximately normal organic loadings and conservative aeration detention times (approximately 5-6 hours mixed-liquor aeration detention time (based on raw flow only) and MLVSS concentrations of 1500-3000 mg/l).
3. Evaluation of the treatment effectiveness and overall process performance of an oxygen aeration system operating at high dissolved oxygen concentrations (7-10 mg/l), high organic loadings (up to 300 lbs BOD/1000 cu ft of mixed-liquor), high MLVSS concentration (4000-5000 mg/l) and very low aeration detention times (1.0-1.5 hours based on raw sewage flow only).
4. Comparison of process treatment effectiveness and economics of an air aeration system operating at relatively high organic loadings, high MLVSS concentrations (2500-3000 mg/l), high aeration rates

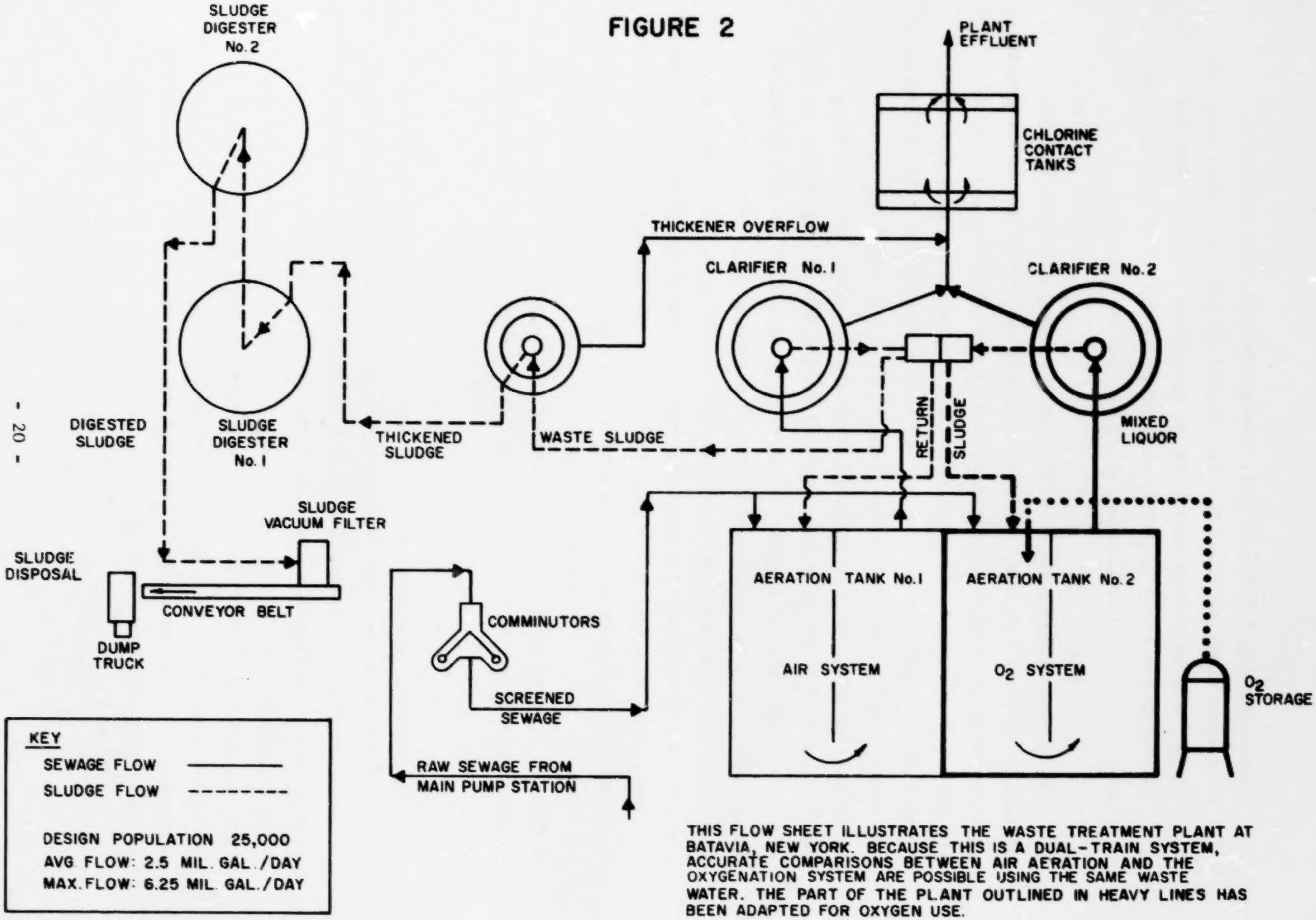
(approximately 3.0-3.5 cf/gal. sewage treated) and treatment detention times of 2.5-3.0 hours (based on raw flow only) with an oxygen aerated system operating at equivalent organic loadings and detention times at MLVSS concentrations of 4000-5000 mg/l and dissolved oxygen concentration controlled at 7-10 mg/l.

The direct comparisons described were possible by simultaneous operation of identical dual train activated sludge systems having completely separate aeration tanks, clarifiers, and sludge return pumps. These studies were conducted using the Batavia, New York Municipal Treatment Plant with the full cooperation and consent of the local city officials and their very capable treatment plant staff. The existing Batavia treatment facility was completed in 1966 and designed for an average daily flow of 2.5 MGD at an average raw sewage BOD of 225 mg/l. As data presented here will illustrate,the average yearly flow is now in excess of 3.0 MGD with flow variations from 1.0-6.0+ MGD. The BOD strength of the incoming wastewater may vary from 100 to 400 mg/l. The wastewater contains a representative mixture of domestic sewage and industrial waste. The wastewater flow was split equally between the air and oxygen aeration systems in comparative cases. A schematic diagram of the Batavia plant is shown in Figure 2. The plant has no primary treatment. Unsettled sewage is treated directly in the aeration system which is designed for contact stabilization operation as well as conventional activated sludge treatment.

The project consisted of four principal phases including the installation of a temporary multistage oxygen aeration system followed by three testing and evaluation phases of operation. The overall operation plan is shown in Figure 3. In the first mode of operation, designated Phase I, all four aeration bays were utilized. Two bays were aerated with air and two with oxygen in a six stage gas contacting system. A schematic diagram of this mode of operation is shown in Figure 4 with indicated average raw sewage flows to each aeration train. In Phase II of operation, the air aeration system was taken out of service along with

WATER POLLUTION CONTROL PLANT, CITY OF BATAVIA, N.Y.
SCHEMATIC FLOW DIAGRAM

FIGURE 2


SLUDGE DIGESTER No. 2
SLUDGE DIGESTER No. 1
DIGESTED SLUDGE
THICKENED SLUDGE
WASTE SLUDGE
THICKENER OVERFLOW
CLARIFIER No. 1
CLARIFIER No. 2
PLANT EFFLUENT
CHLORINE CONTACT TANKS
RETURN SLUDGE
MIXED LIQUOR
SLUDGE VACUUM FILTER
SLUDGE DISPOSAL
CONVEYOR BELT
DUMP TRUCK
COMMINUTORS
SCREENED SEWAGE
RAW SEWAGE FROM MAIN PUMP STATION
AERATION TANK No. 1
AIR SYSTEM
AERATION TANK No. 2
O_2 SYSTEM
O_2 STORAGE
KEY
SEWAGE FLOW
SLUDGE FLOW
DESIGN POPULATION 25,000
AVG. FLOW: 2.5 MIL. GAL./DAY
MAX. FLOW: 6.25 MIL. GAL./DAY

THIS FLOW SHEET ILLUSTRATES THE WASTE TREATMENT PLANT AT BATAVIA, NEW YORK. BECAUSE THIS IS A DUAL-TRAIN SYSTEM, ACCURATE COMPARISONS BETWEEN AIR AERATION AND THE OXYGENATION SYSTEM ARE POSSIBLE USING THE SAME WASTE WATER. THE PART OF THE PLANT OUTLINED IN HEAVY LINES HAS BEEN ADAPTED FOR OXYGEN USE.

FIGURE 3

PROJECT OPERATION PLAN

PROJECT MONTH

AIR SYSTEM	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
	Normal Operation	Diffusers and Instrumentation installed, Piping modifications made, Bulkhead door installed.				Operation to Acclimate Sludge	Phase I		System Shutdown	Not in Service	Phase III		Removal of temporary equipment	Final Report Preparation	

OXYGEN SYSTEM	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
	Equipment Designed and Ordered	Equipment and Instrumentation installed. Plant modifications carried out.				Operation to Acclimate Sludge	Phase I		System Conversion	Phase II	Phase III		System Conversion to air	Final Report Preparation	

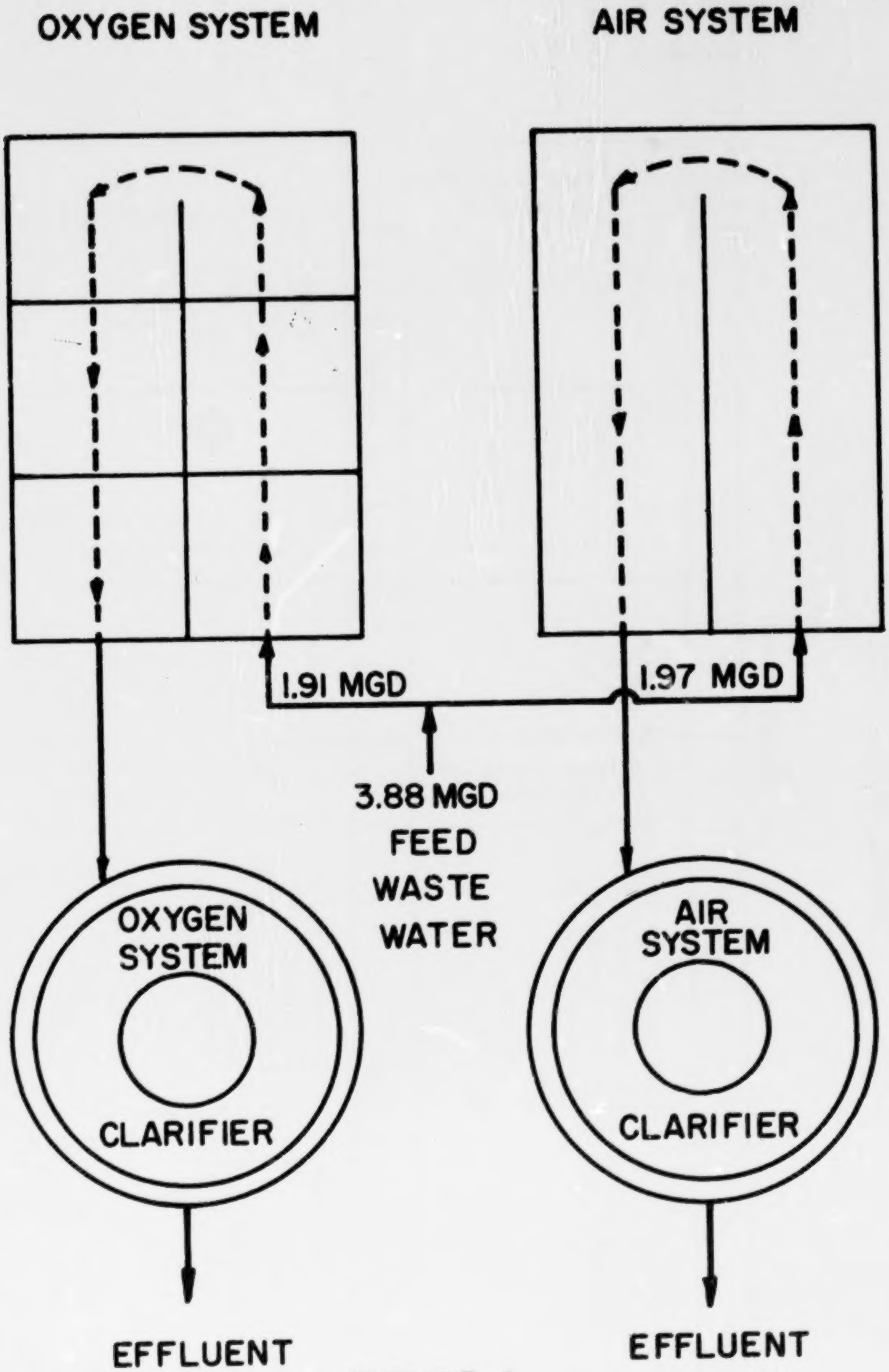

OXYGEN SYSTEM
AIR SYSTEM
1.91 MGD
1.97 MGD
3.88 MGD
FEED
WASTE
WATER
OXYGEN
SYSTEM
CLARIFIER
AIR
SYSTEM
CLARIFIER
EFFLUENT
EFFLUENT

FIGURE 4

BATAVIA TREATMENT SYSTEM
PHASE * I OPERATION

one bay or three stages of the oxygen aeration system. The remaining three stage oxygen aeration system using 1/4 of the available aeration tankage was then used to treat the entire raw sewage flow. This system configuration is shown diagramatically in Figure 5. In the final evaluation, Phase III, one bay each of the air and oxygen systems were used to treat approximately equivalent raw sewage flows. The resulting comparison was that of a three stage gas-liquid oxygen contacting system with a conventional high rate air aerated activated sludge system. This configuration shown in Figure 6 indicates the addition of liquid staging baffles to the air system. This staging baffle addition was done to permit identical staging of mixed-liquor for both the air and oxygen systems. Under these conditions any dissimilarities in the performance of the two systems attributable to liquid staging were eliminated.

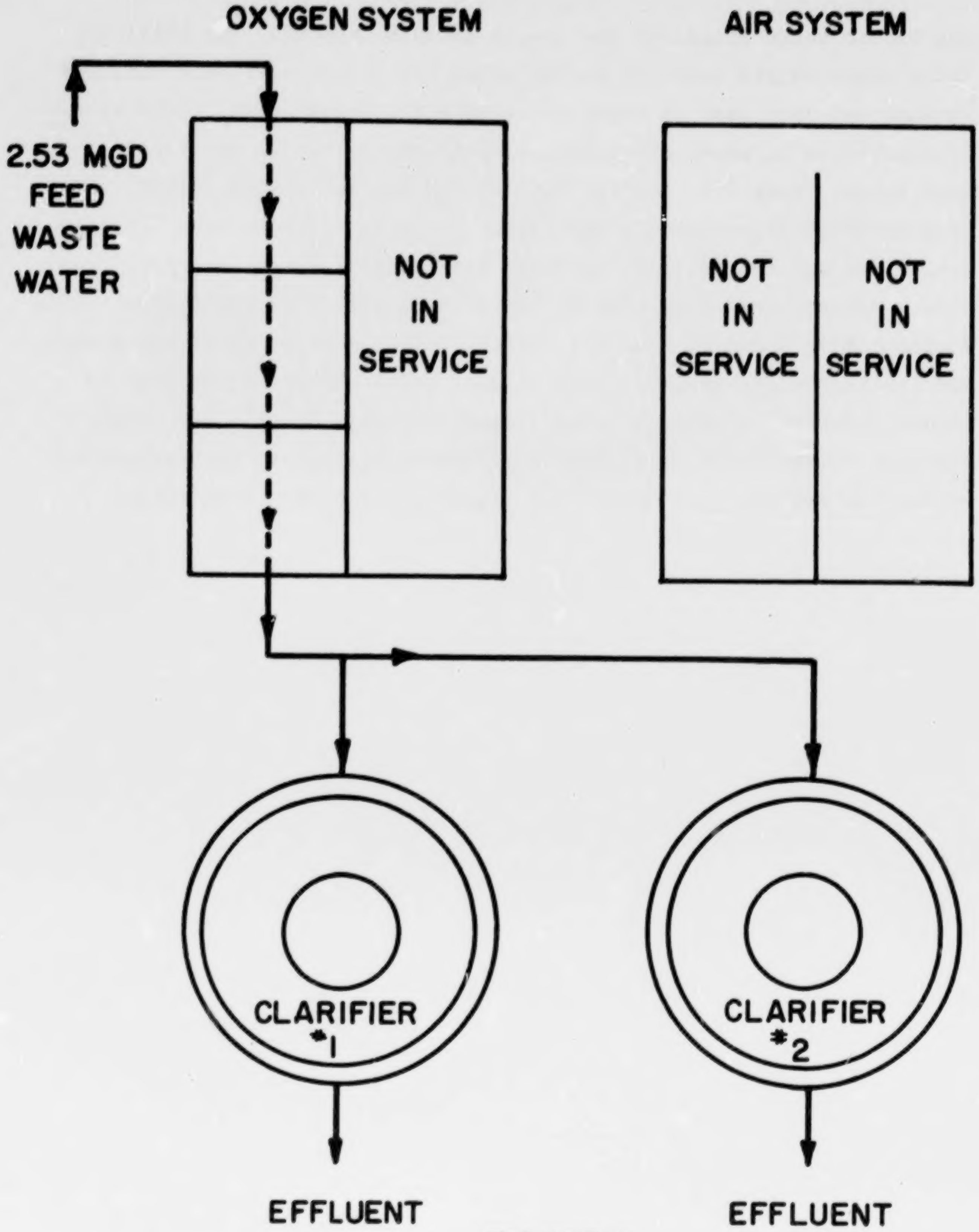

OXYGEN SYSTEM
AIR SYSTEM
2.53 MGD FEED WASTE WATER
NOT IN SERVICE
NOT IN SERVICE
NOT IN SERVICE
CLARIFIER *1
CLARIFIER *2
EFFLUENT
EFFLUENT

FIGURE 5

BATAVIA TREATMENT SYSTEM
PHASE *2 OPERATION

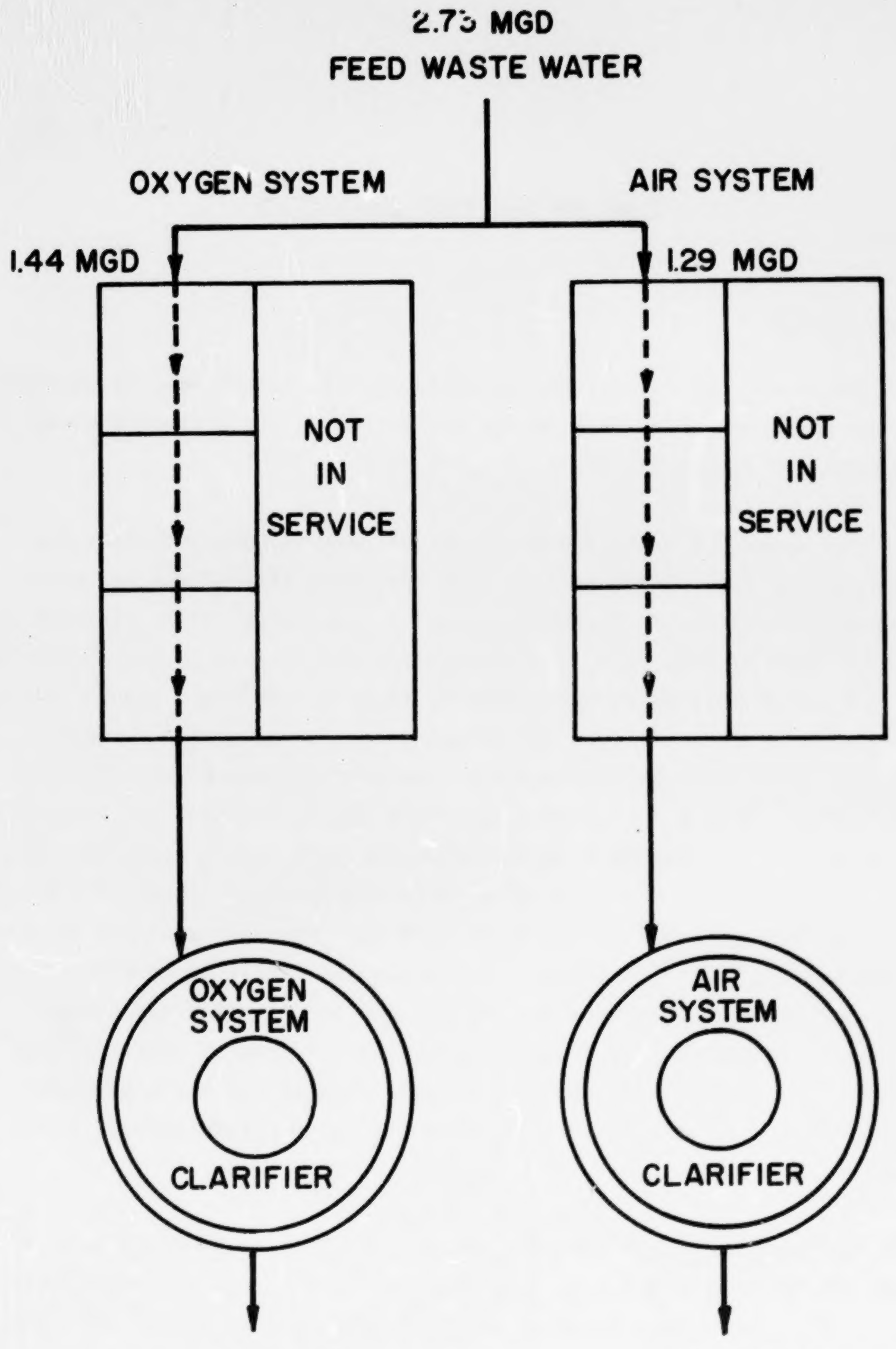

2.73 MGD
FEED WASTE WATER
OXYGEN SYSTEM
AIR SYSTEM
1.44 MGD
1.29 MGD
NOT IN SERVICE
NOT IN SERVICE
OXYGEN SYSTEM
CLARIFIER
AIR SYSTEM
CLARIFIER
EFFLUENT
EFFLUENT

FIGURE 6

BATAVIA TREATMENT SYSTEM
PHASE # 3 OPERATION

PLANT AND EQUIPMENT DESCRIPTION

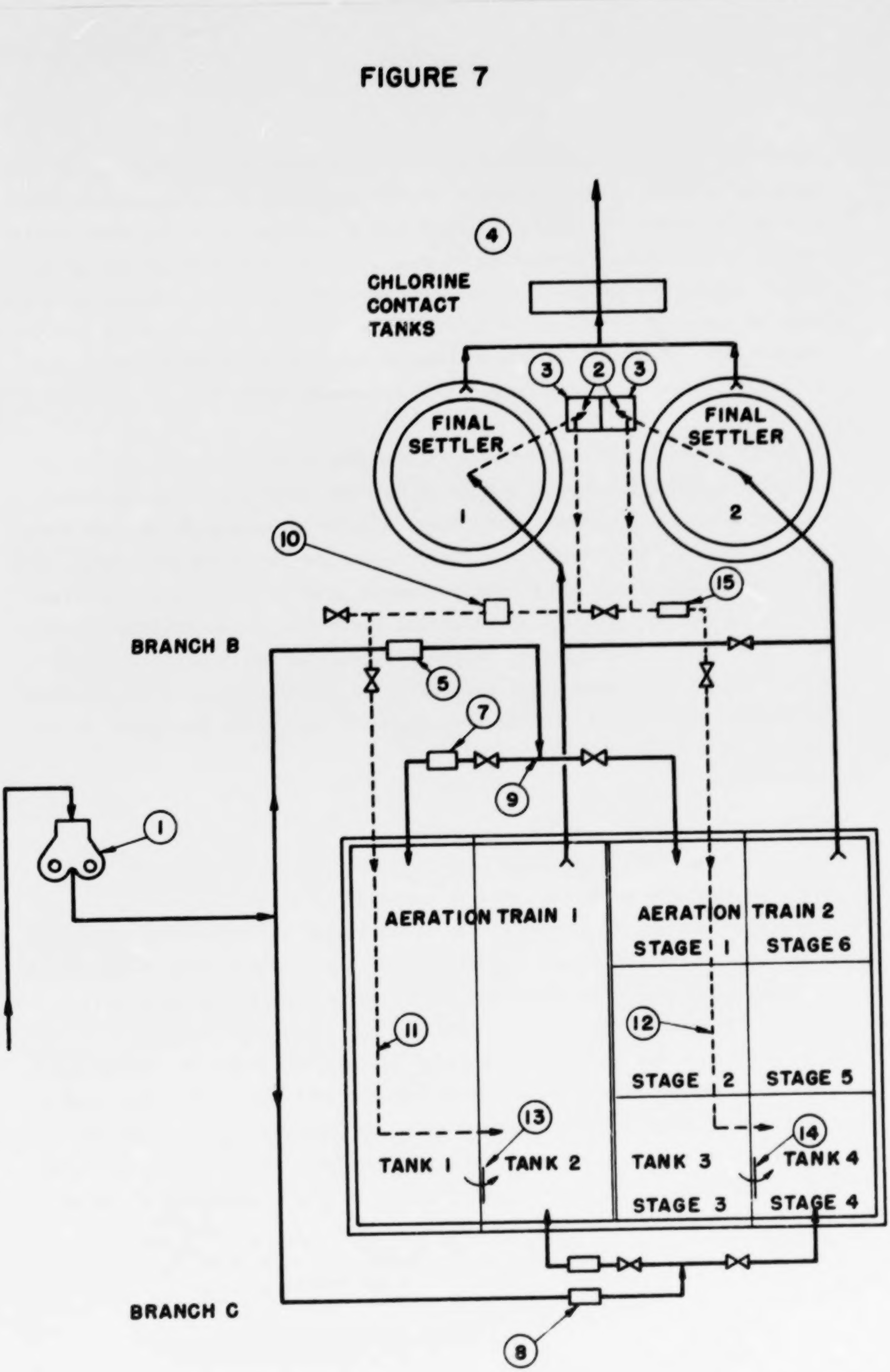
Plant Layout

The plant description for purposes of this report will be confined to the aeration and/or oxygenation facilities in the activated sludge secondary treatment process.

The water pollution control plant at Batavia, New York uses the diffused air aerated activated sludge treatment process for pollutant removal. The plant is designed for a 2.5 MGD average daily flow and a maximum flow of 6.25 MGD. Raw sewage from the Batavia collection system enters a pump station located remotely from the treatment plant. Grit and debris are removed from the sewage before it is pumped through a 14-inch force main into two parallel comminutors located at the treatment plant. (Point 1 on Figure 7). From the comminutors the sewage flows by gravity through a 30 inch diameter main to the aeration tanks. As seen in Figure 7, four aeration tanks are used and connected hydraulically to form two parallel treatment systems. The piping of the aeration tanks is such that both conventional activated sludge or contact stabilization treatment processes can be employed by directing raw sewage through pipe branch B or C respectively. The volume of each aeration train (two tanks) is 325,200 gallons resulting in 6.2 hours of mixed-liquor detention time based on 2.5 MGD feed flow. Each tank is 23 ft. wide and 64 ft. in length.

The circular final settlers are designed for an overflow rate of 1000 gal/ft^2/day at 2.5 MGD. Settled activated sludge is pumped from the settlers by two 6-inch diameter air lift pumps (2) to a sink (3) from

FIGURE 7


④
CHLORINE
CONTACT
TANKS
③
②
③
FINAL
SETTLER
1
FINAL
SETTLER
2
⑩
⑮
BRANCH B
⑤
⑦
⑨
①
AERATION TRAIN 1
AERATION TRAIN 2
STAGE 1
STAGE 6
⑪
⑫
STAGE 2
STAGE 5
⑬
⑭
TANK 1
TANK 2
TANK 3
TANK 4
STAGE 3
STAGE 4
BRANCH C
⑧

which the return activated sludge flows by gravity to the feed end of the aeration trains. Waste activated sludge also flows by gravity from sink (3) to the sludge thickener. Return sludge pumping rates for each settler can be independently controlled to meet process conditions at any given time. Piping of the final settlers is such that both may operate with the same or separate aeration tanks, a feature necessary for parallel process evaluation. The final treatment step at Batavia is chlorination of the effluent before it is discharged into Tonawanda Creek (4).

The aeration rate at Batavia is normally 1.83 cubic feet of air per gallon of sewage treated at design flow. Air injection is accomplished through knee joint, swing arm diffuser assemblies employing Chicago Pump, Inc. flat disk, coarse bubble diffuser heads as gas injection units.* A number of modifications to the treatment plant were necessary to adapt to a separate but parallel simultaneous operation of two different treatment processes. Furthermore, it was necessary to build in flexibility beyond normal treatment plant capability to successfully meet the process parameters specified in the various phases of operation described in the Experimental Program Plan.

Aeration Tank Modifications

The multistage high purity oxygenation process used in this study requires that the aeration tanks be enclosed by a gas tight cover. Also, both gas and activated sludge mixed-liquor are cocurrently staged. These conditions can be met readily during new aeration tank construction in which both staging baffles and cover may be made an integral structural part of the tanks. At Batavia, however, the tanks were existing ones and the conversion was a temporary measure designed for ease of installation and final removal upon completion of the contract work. For this reason, steel rather than concrete was chosen as the material for baffles and covers. A total of six steel I-beam columns were used for structural support of each of the two staging baffles placed in each tank. These columns were spaced at 38 inch center distances to form a skeleton across the tank. Each column was anchored to the tank bottom by concrete anchors and bolted to angles traversing the tank at the top. Liquid and gas staging was achieved by gasketing and bolting 1/8" thick steel sheets to the support columns. The gas seal between the tank walls and the steel baffle was achieved by contour fitting and gasketing a flanged steel sheet against the concrete walls. The gasketing used was an oxygen compatible material. The gasketed flange was held against the concrete wall by anchors spaced 6 inches apart. Two openings shown in Figure 8 for interstage liquid flow were located on the tank bottom. Each of these openings is 38 inches high by 34 inches wide. The size of these openings was sufficient to pass liquid from stage to stage under low hydraulic head while small enough so that each stage could be considered fully separated from the next relative to mixing and oxygenation.

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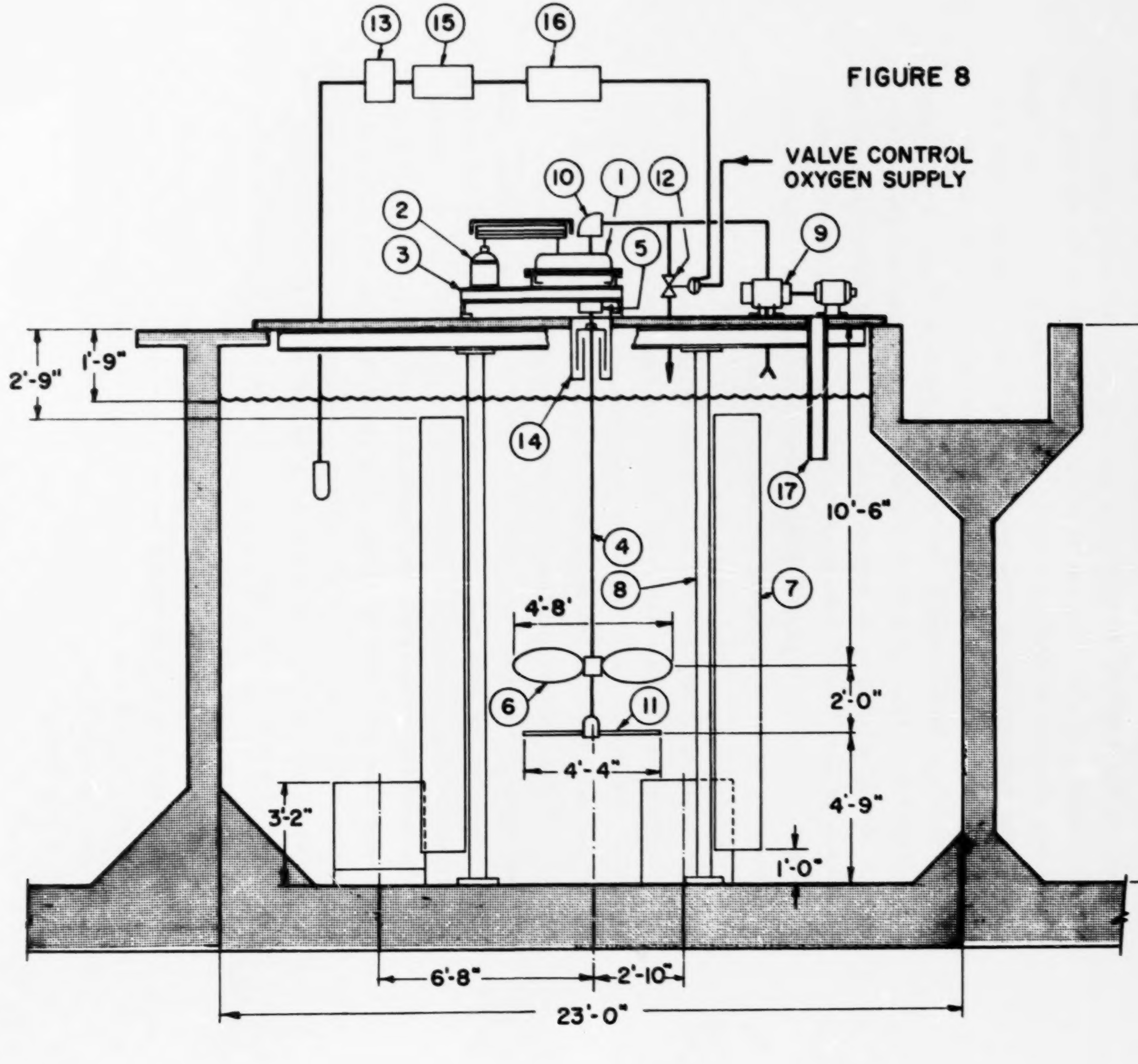
* Mention of a proprietary device does not imply endorsement by the Federal Water Quality Administration.

The cover structure for the oxygenation system was designed for two main functions, to contain the oxygenation gas over the mixed-liquor at a pressure not exceeding 12 inches of water column and to support the mixing and gas recirculation equipment. Structurally the cover was supported by two I-beams (Figure 8) spanning the tank. Each beam in turn was supported by two pipe columns anchored to the tank bottom. The I-beam elevation was such that its upper flange was flush with the concrete wall tops allowing for smooth transition of cover steel from the I-beam over the concrete. Pre-fabricated and reinforced 1/8 inch thick steel sheets were used for cover material, each sheet spanning the tank and overlapping the concrete. A gas tight seal between the concrete walls and the steel covers was maintained by an oxygen compatible gasketing material. Each cover sheet was compressed against the gaskets by bolted down clips along all edges of the cover sheet. Access to the oxygenation stages was provided by a 20-inch diameter manway located in each cover. The manway covers served as safety relief gas vents in the event of overpressurization of the system. Relief settings of eight inches of water column gauge pressure were maintained by appropriate water filled seals in each manway cover.

FIGURE 8

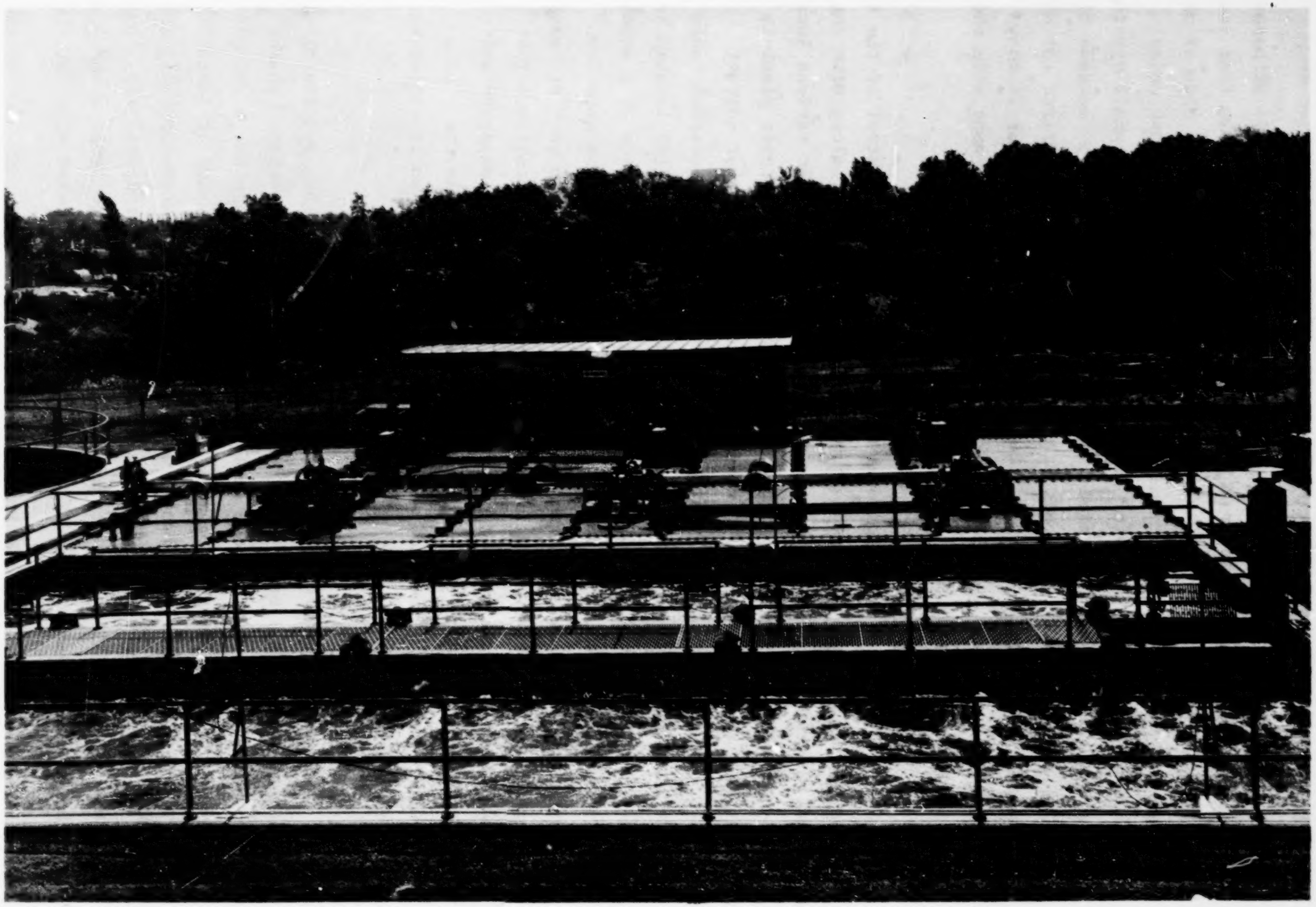

13
15
16
VALVE CONTROL
OXYGEN SUPPLY
10
1
12
9
2
3
5
2'-9"
1'-9"
14
17
10'-6"
17'-3
4
8
7
4'-8'
6
11
2'-0"
4'-4"
3'-2"
4'-9"
1'-0"
6'-8"
2'-10"
23'-0"

Sample ports (Item 17, Figure 8) for mixed-liquor sample withdrawal were provided in each stage. These sample ports extended two feet into the liquor to prevent gas loss during sampling. Interstage flow of oxygen gas was through a 3-inch diameter tube located above the tank cover. Normally interstage transfer of oxygen would be through a small opening across the gas staging baffles beneath the tank covers, but because of the need for gas stream monitoring this stream was exteriorized for convenience. An overall view of the cover and equipment layout is shown in Figure 9. The air aeration tanks are shown in the foreground with the six stage oxygenation unit in the background.

The need for complete separation and independent control of the two parallel processes required a number of piping changes, along with the addition of liquid flow and air flow metering stations at various locations within the plant. As pointed out earlier, the Batavia plant is piped to operate with the conventional activated sludge or contact stabilization process. In either case,raw feed to the aeration tanks is measured by one Fisher Porter magnetic flow meter (5)(8) located in each feed line branch (B) and (C) (Figure 7). Installation of a second Fisher Porter 14-inch diameter magnetic flow meter (7) was made down stream of the flow splitting tee (9). Totalized flow to each process was then computed from differences on the integrator totalizers for each metering station. It was desirable to completely separate the feed piping to each process, however, space available in the piping tunnel and cost associated with this extensive piping modification was prohibitive.

Return activated sludge piping in the plant did not require alterations for Phase I operation, however, subsequent operational phases required that return sludge be fed to oxygenation stage No. 4 and also into air train aeration bay No. 2. This was accomplished by installing (see Figure 7) 16-inch diameter sludge return pipe extensions (11) and (12) into the aeration and oxygenation bays 1 and 3 respectively. During Phases II and III operation, waste was fed to aeration tanks 2 and 4 only through feed branch (C), while tanks 1 and 3 were blocked off from the

FIGURE 9 OVERALL VIEW OF AIR AND OXYGENATION SYSTEMS AT BATAVIA



hydraulic path by installing liquid bulkheads (13)(14) in the interconnecting passage between the aeration tanks. During Phase II and III operation, return sludge was fed through pipe (11) and/or (12) into tanks 2 and/or 4 by connecting lines (11) and (12) to the liquid bulkheads (13) and (14) shown in Figure 7.

Adequate return sludge pumping and metering capability appeared to be available in the plant for Phase I operation. Subsequent experience in Phase I indicated the requirement that two new 10-inch diameter air-lift sludge return pumps be installed in lieu of the existing 6-inch diameter air lift pumps. Also, during Phase II operation when the total plant recycle sludge was returned to bay No. 4, it was necessary to install a new Fisher Porter magnetic flow meter (10-inch diameter) to replace an existing 4-inch diameter meter located at (15). No change was necessary for the 4-inch recycle sludge meter of the air system located at (10) for either Phase I or III operation.

Oxygenation Equipment

The multistage activated sludge oxygenation process used depends upon separate mechanical equipment components for liquid mixing, oxygen compression and dissolution. Both liquid mixing and oxygen dissolution energy requirements for each stage in the system will vary considerably. Therefore, each stage is equipped with an independent mixer-oxygenator combination designed to provide only the required level of mixing and oxygenation for that particular stage. This arrangement results in very efficient oxygen transfer energy utilization through judicious matching of efficient mixing and oxygen dissolution equipment to the requirements of each stage throughout the multistage contacting system.

A schematic diagram of one of the contacting stages used at Batavia is shown in Figure 8. This diagram shows an oxygenator configuration where large fluctuations in organic and hydraulic loadings are expected, hence an automatic dissolved oxygen level feed back control system is used. This control, which operates by varying the gas recirculation rate within the stage, may be provided where desirable.

The oxygenation equipment used at Batavia was specifically custom designed to enable efficient operation under the wide range of process conditions evaluated in the course of the contract work. The mixer drive (1) was a Link-Belt, shaft mounted speed reducer,* (size 307D20) providing a double gear reduction of 20:1. This speed reducer was designed to mount directly onto a shaft to be driven with the shaft passing through the speed reducer. This feature was necessary in order to inject oxygen gas into the propeller shaft (4) at point (10) as shown in Figure 8.

The speed reducer was driven by a 1725 rpm electric motor (2) through four "B" size v-belts and shieves. The output shaft (hereafter called propeller shaft) speed, when a 1:1 shieve ratio was used, was 86.25 rpm. Propeller shaft speeds of 62.5 and 74.5 rpm were also provided through selective shieve ratios. The speed reducer (1), electric motor (2), and propeller shaft (4) combinations were mounted on a frame (3) designed to integrate and support the complete oxygenator assembly. The propeller shaft (4) was a hollow carbon steel tube, 3-1/2 inches in outside diameter with a one half inch thick wall. This shaft supported a three bladed, 56 inch diameter marine propeller (6) suspended 126 inches below the tank cover. Since the overhanging loads at this shaft extension became quite excessive for the speed reducer to withstand, a steady bearing (5) was designed into the support frame and was located approximately 20 inches below the speed reducer top bearing.

The gas injection sparger device (11) was attached to the propeller shaft 24 inches below the propeller heel center line. The sparger device consisted of a hollow center hub which was threaded to the propeller shaft' and from which 1-1/4 inch diameter pipes extended radially to form the sparger arms (eight arms per assembly). Each arm as it extends outward from the hub was swept back slightly. This swept back design was sufficient to shed rags or other debris which might collect on the arms as the sparger rotates. To minimize drag load on the arms, each arm was tapered by uniformly flattening the arm from 1-1/4 inch outside diameter at the hub to a 1/2 inch thick, round edged, flat cross section at the tip. Each arm was drilled through (top and bottom) with 306, 1/8 inch diameter holes resulting in 2448 holes per sparger. Oxygen from the gas space in each stage was fed to the sparger gas-injection unit by a positive displacement, three lobe, rotary type compressor (9) through a rotary gas seal (10) and down the propeller shaft (4). The compressor used was manufactured by MGD Pneumatic, Inc.* using hard coated aluminum construction for housing and rotors. Gas leakage to the atmosphere past the rotor shafts is sealed by high pressure carbon face seal combinations. The compressor used in Stage No. 1 (Phase I operation) in Batavia was Model 4012 delivering 280 cfm/NTP at 1850 rpm, with 14.7 psia suction and 20.7 psia discharge pressures. The compressor was driven by a 1750 RPM electric motor through a 1.06:1 V-belt sheave ratio. In Phase I operation,the Stage No. 2 compressor delivered 180 cfm/NTP while those on Stages No. 3, 4, 5 and 6 delivered 85 cfm/NTP. In Phases II and III of the operation (3 stage system used),the first stage compressor delivered 180 cfm/NTP while the second and third stage units delivered 85 cfm/NTP each.

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* Mention of a proprietary device does not imply endorsement by the Federal Water Quality Administration.

The entire Stage No. 1 compressor output was seldom required to maintain adequate dissolved oxygen levels in the mixed-liquor, therefore, an automatic D.O. feedback control and control valve (12) was installed in a bypass arrangement to regulate flow of oxygen to the sparger. The feedback control sensing device was a Union Carbide Corporation, Model 1101 dissolved oxygen probe (13) and analyzer* giving a 0 to 50 millivolt linearized output signal. This signal was transmitted to a millivolt-ampere converter (15) having a 1 to 5 ma output (Transmatron Inc. Model 330T)*. The converted signal (1-5 ma) was then used by an indicating controller (16) (Robertshaw Controls Co. Model 321-A1-S2)* with total set point indication, manual set point, process variable and valve opening

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indication. The 1-5 ma signal was also fed to a 2 inch pipe size automatic diaphragm valve positioner. Proportional conversion of the 1-5 ma signal to pneumatic activation of the positioner controlled the amount of bypass gas and consequently the degree of oxygen injection to the mixed-liquor in Stage 1. This dissolved oxygen control system, used for the work described here, was necessary to meet the specific objectives of the evaluation program described in the contract. In permanent installations this option would be available, using a variable speed blower, but is not in general a necessary control system requirement for a multistage oxygenation unit.

An item not previously described is the anti-swirl baffles (7) in the oxygenation stages. Four baffles measuring 24 inches wide by 159 inches long were attached to the tank cover and oxygenator support columns (8) and extended radially outward on a line drawn from the propeller shaft and through the aerator support columns. These baffles were installed to prevent excessive vortexing and swirl by the agitator.

The gas seal (14) around the propeller shaft was achieved by two overlapping cups or cylinders, the lower cup being attached to the cover plate and containing the seal water, the upper cup being attached to the shaft and sealed to it by an "O"-ring. No operational problems were experienced with this simple hydraulic seal arrangement.

To prevent instantaneous changes in tank liquid level, and thus changes in pressure, a mixed-liquor effluent overflow weir was installed. This weir served also to prevent bubble entrainment in the effluent liquor as might occur if the mixed liquor effluent line opening were submerged.

An overall view of the oxygenation system has been shown in Figure 9, and a close up of a single oxygenator drive, compressor and automatic control valve arrangement is shown in Figure 10. It should be pointed out that these gas contacting units were custom designed to enable operation under a wide variety of process conditions and were highly instrumented. The device in Figure 10, therefore, does not accurately reflect the type of equipment which would be employed in a permanent installation.

Aeration System Modifications

Figure 11 shows a diagram of the air aeration piping system at the Batavia plant. Since aeration tanks 3 and 4 were used for the oxygenation system, the air header (2) valve (3) was closed and this leg of the piping was not used. Compressed air from the three blowers was fed through an 18-inch line to the 14-inch header (1) for diffuser banks (8) in aeration tanks 1 and 2 as required in the various phases of operation. The air blowers were used alternatively with the 225 HP gas engine (blower 1) usually being used during the day and the 125 HP (blower 2) electric machine being used during the night. To increase the aeration rate capacity in tanks 1 and 2, 100 (50% increase) additional diffusers were added to the existing banks. Tapering of the aeration rate was accomplished by adjustment of the valves (4) located on each diffuser bank swing arm header.

The amount of excess air vented to the atmosphere was regulated manually by adjustment of the valve shown as item (5). This excess air flow was measured by an orifice flow meter (6), the differential pressure being recorded continuously (7). Automatic control of the air wasting rate was established using a dissolved oxygen feed back control system (8)(9)(10) identical to that described for the oxygenation system, Stage 1. This was done to attempt to meet day-night changes in oxygen demand so that a fairly constant D.O. concentration could be maintained throughout the air aeration system mixed-liquor. In general, it was necessary to use almost the entire air blower output to maintain the desired D.O. concentration. As a result, only a small fraction of the total blower output was wasted. Measurement of the blower output less the quantity of air wasted permitted calculation of the net compression power required to operate air tanks 1 and 2.

FIGURE 10 VIEW OF STAGE NO. 1 OXYGENATION SYSTEM AERATOR, BLOWER AND TURBINE DRIVE

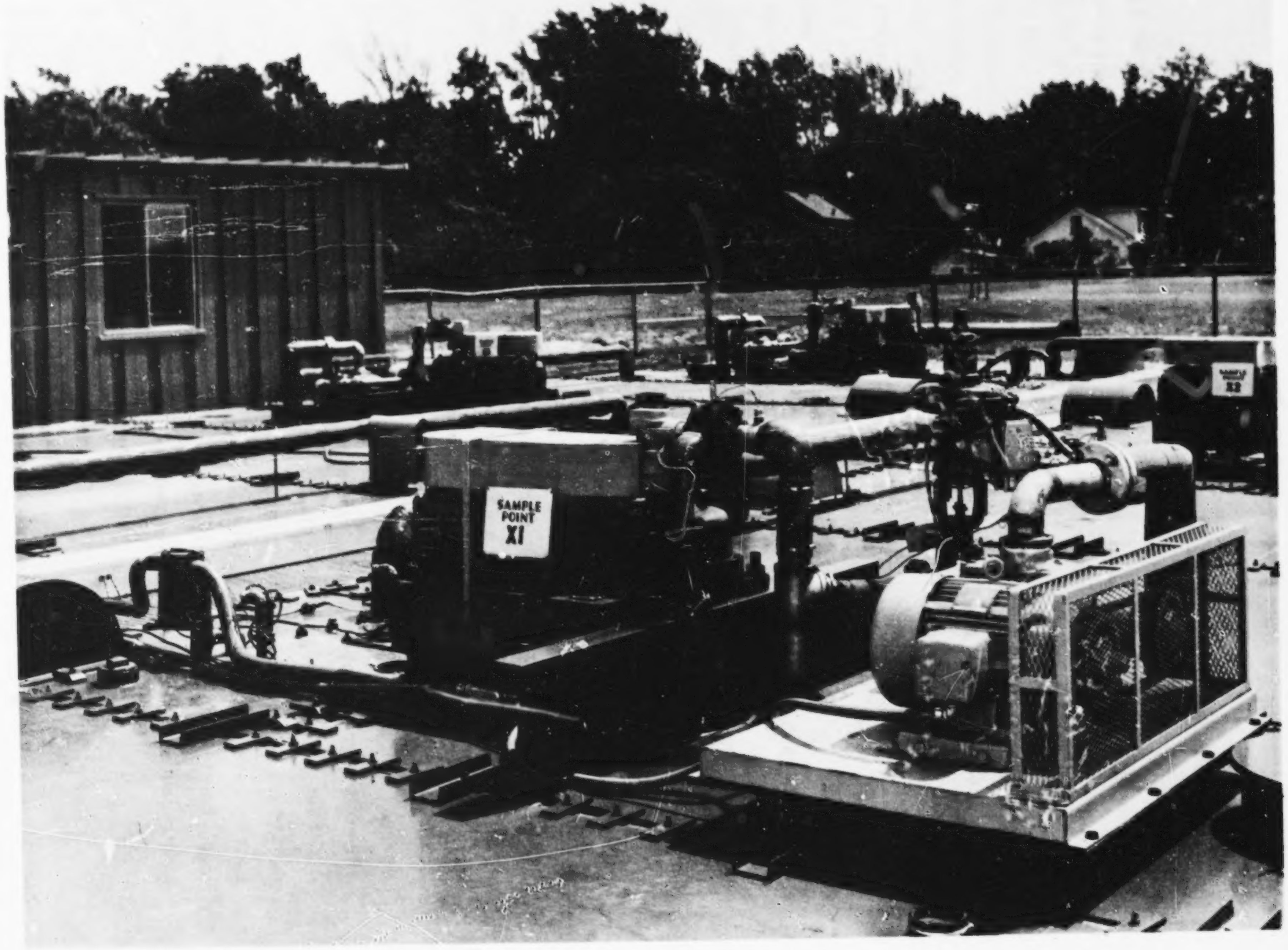
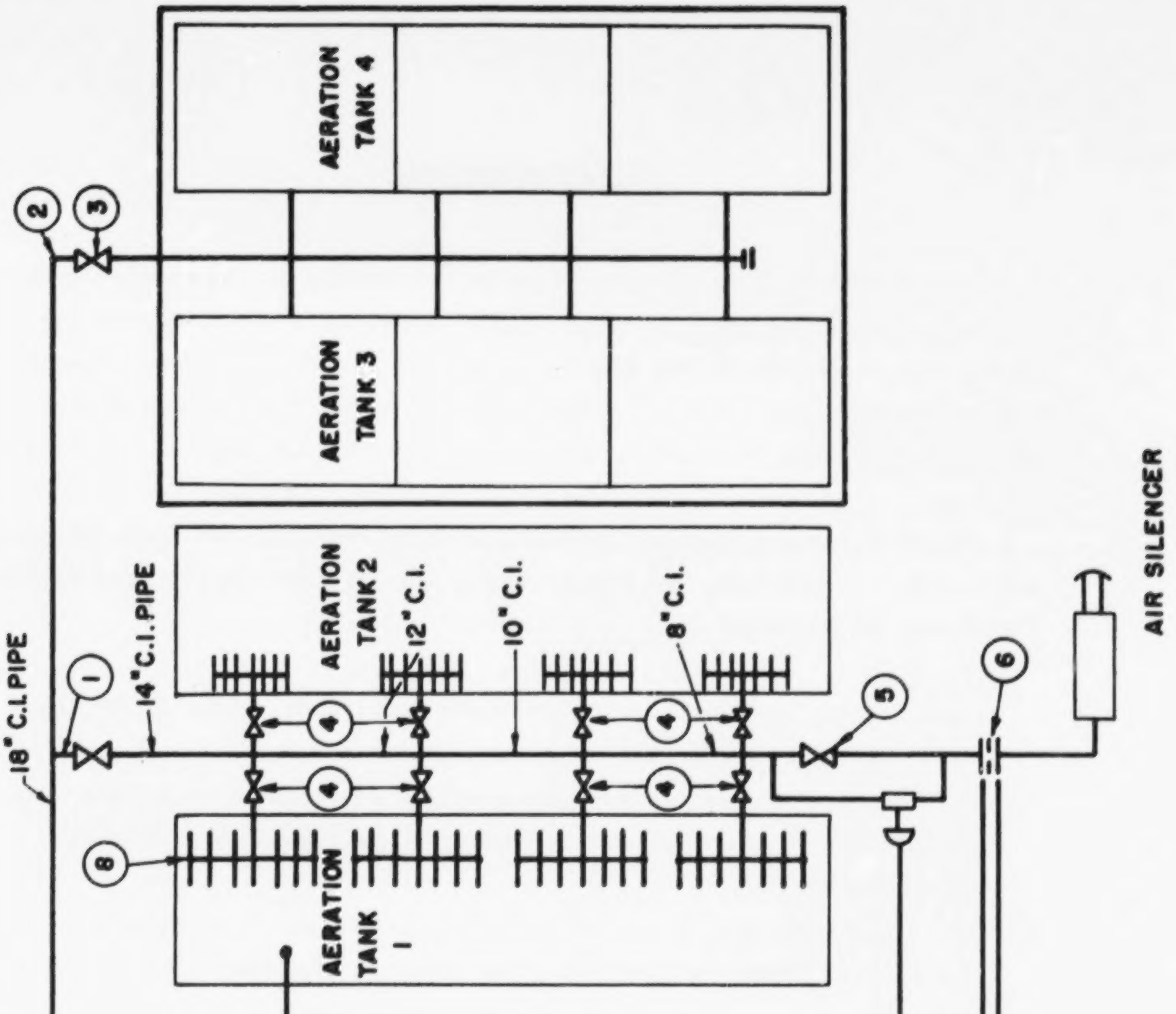

SAMPLE POINT XI

FIGURE II


18" C.I PIPE
14" C.I. PIPE
①
②
③
AERATION TANK 4
AERATION TANK 3
AERATION TANK 2
12" C.I.
10" C.I.
8" C.I.
④
④
④
④
⑤
⑥
AIR SILENCER
⑧
AERATION TANK 1
BLOWER 3
3980 CFM
870 RPM
150 HP ELECTRIC MOTOR
BLOWER 2
3180 CFM
1200 RPM
125 HP ELECTRIC MOTOR
BLOWER 1
3275 CFM
1200 RPM
225 HP GAS ENGINE
⑩
⑨
⑧
⑦

EVALUATION METHODS

Concurrent with the installation of the oxygen aeration system at Batavia, a combination laboratory control room and process monitoring center was installed in the plant operation building. It was intended that continuous recording of nearly all aspects of the systems operation would be conducted throughout the eight month test period.

Several basic groups of information were required for performance and economic evaluation and comparison of the air and oxygenation systems. These were as follows:

1. Raw sewage, sludge recycle, and waste activated sludge flows for both air and oxygenation systems.
2. Suspended and volatile suspended solids concentrations in raw sewage, recycle sludge, mixed-liquors, and clarifier effluents for each system.
3. Physical characteristics of the mixed-liquor as determined by sludge volume index and initial settling velocity measurements in each system.
4. Chemical and biochemical characteristics of the feed raw sewage, mixed-liquor and final settler effluent in each system as revealed by BOD, COD, ammonia nitrogen (NH_3-N), nitrate nitrogen (NO_2-N), nitrate nitrogen (NO_3-N), total Kjeldahl nitrogen (TKN) and total phosphorous (TP) determinations in each case.
5. Dissolved oxygen concentrations at several points in both the air and oxygenation systems as continuously monitored and recorded.
6. Gas phase oxygen composition in the gas space of each of the stages in the oxygenation system and in the vent gas as continuously monitored and recorded.
7. Overall oxygen utilization measurement in the oxygenation system

by measurement of feed and exhaust gas volumes and measurement of the rate of aeration (cf air/gal. waste treated) in the air system.

8. Measurement of actual power consumed for gas recirculation and liquid mixing in the oxygenation system and of power required for air compression in the air aerated system.
9. Measurement of all temperatures and pressures necessary in various parametric calculations.
10. Measurement of gas recirculation rates within stages of the oxygenation system.

To monitor these various groups of parameters, appropriate metering and recording equipment was installed where possible, supplementing the activities of on-site technicians available seven days per week to carry out the experimental evaluation.

Sampling Methods and Frequency

Continuous composited sampling of influent raw wastewater and final settler effluents was carried out throughout the course of the work. These composite samples were taken for two periods of time each day including 11:00 p.m. (previous day) to 10:00 a.m. and 10:00 a.m. to 11:00 p.m. with the various analyses being performed individually in each instance. These time periods were chosen to compensate for differences in day versus night organic and hydraulic loads to the plant. It was not possible to flow proportion the composite sampling system so that separation of the day into two sampling periods substantially reduced inherent errors while providing the basis for additional detailed information. Liquid and gas flow meter integrators were read to correspond to the described sampling intervals. In summary data tabulations to be reported here, night time and separate day time values are reported for all parameters. The 24 hour mean value is the flow weighted average of day and night operating values in each instance.

Grab samples of mixed-liquor and recycle sludge were taken for each system three times daily and the combination of these three samples, in each instance, was used as the composite for TSS (Total Suspended Solids) and VSS (Volatile Suspended Solids) analysis. Where mixed-liquor from a multistage system was evaluated, each liquid stage was sampled by the foregoing protocol, yielding 3 to 6 mixed-liquor solids determinations per day, each being the combination of three grab samples.

Where SVI or settling velocity of the mixed-liquor activated sludge was to be measured, special grab samples were taken from each aeration stage or tank and were immediately evaluated. Supernatant fractions of these samples (unfiltered, settled only) were analyzed for BOD, COD, NH_3-N, NO_2-N, NO_3-N, TP, and TKN content periodically.

When sludge wasting was done for either system (usually a 4-5 hour period/day), special grab samples of recycle sludge were taken during this period for VSS and TSS concentration measurement. This permitted a more accurate estimation of the quantity of dry solids wasted from a given aeration train. This sampling of recycle sludge during wasting, as well as the previously described three times per day sampling, was necessitated by occurrence of hour to hour variation in the recycle sludge solids concentration. This condition occurred as a result of the manual operation of recycle sludge pumps (air lifts) rather than a more desirable flow proportioned control. Hour to hour changes in raw sewage flow were not always compensated by manual changes in the recycle flow rate. The resulting variation in the ratio of recycle sludge to raw sewage flow commensurately affected changes in recycle sludge solids concentrations. These changes were particularly evident during the early morning hours each day when no plant operator was on duty.

At all times the oxygenation and air systems at Batavia were operated by regular plant personnel. Linde technicians were on-site only 8 hours per day for five days each week to conduct analysis and monitor the systems. At all other times, weekends and evenings, off duty plant operators were retained to collect samples, conduct certain analyses and monitor the operation. No regular plant operating personnel or Linde technicians were on duty from 11:00 p.m. each night to 7:00 a.m. each morning. This was a period of unattended operation. Late night samples were taken by off-time operators in most cases.

In Tables 4, 5, and 6 are shown the various analytical measurements made for both air and oxygenation systems on feed raw waste water, mixed-liquor, recycle sludge, and final clarifier effluent. Also indicated is the frequency of sampling and analysis.

TABLE 4

PHASE I OPERATION - WEEKLY SCHEDULE OF ANALYSES

Analysis	Common Feed Wastewater	Mixed-Liquor						Recycle Sludge		Clarifier Effluent	
		Air System			Oxygen System						
		Inlet	Middle	Outlet	No. 1	No. 4	No. 6	Air Clarifier	Oxygen Clarifier	Air System	Oxygen System
pH	C-7 G-3	G-3	-	G-3	G-3	G-3	G-3	-	-	C-7 G-3	C-7 G-3
Turbidity	-	-	-	-	-	-	-	-	-	C-7	C-7
TSS	C-7	K-7 G-3	K-7 G-3	K-7 G-3	K-7 G-3	K-7 G-3	K-7 G-3	K-7 G-7	K-7 G-7	C-7	C-7
VSS	C-7	K-7	K-7	K-7	K-7	K-7	K-7	K-7 G-7	K-7 G-7	-	-
Settleability	-	G-3	G-3	G-3	G-3	G-3	G-3	-	-	-	-
Settling Rate & SVI	-	G-3	G-3	G-3	G-3	G-3	G-3	-	-	-	-
BOD	C-7 G-3	G-3	G-3	G-3	G-3	G-3	G-3	-	-	C-7 G-3	C-7 G-3
COD, TP , TKN, NH_3-N NO_2-N, and NO_3-N	C-7	G-3	G-3	G-3	G-3	G-3	G-3	-	-	C-7	C-7

LEGEND - C - Two continuous composites daily (Night and Day Periods).
G - One grab-sample analyzed daily.
K - Three grab-samples daily at equal intervals (8 hours), combined and analyzed as one.
Number Following Letter - Indicates number of days each week the sample is taken for analysis.

TABLE 5

PHASE II OPERATION - WEEKLY SCHEDULE OF ANALYSIS

Analysis	Feed Wastewater	Mixed-Liquor Oxygen System Stage 1	Stage 2	Stage 3	Recycle Sludge	Combined Clarifier Effluents
pH	C-7 G-3	G-3	G-3	G-3	-	C-7 G-3
Turbidity	-				-	C-7
TSS	C-7	K-7 G-3	K-7 G-3	K-7 G-3	K-7 G-7	C-7
VSS	C-7	K-7	K-7	K-7	K-7 G-7	-
Settleability	-	G-3	G-3	G-3	-	-
Settling Rate & SVI	-	G-3	G-3	G-3	-	-
BOD	C-7 G-3	G-3	G-3	G-3	-	C-7 G-3
COD, TP, TKN, NH_3-N NO_2-N, and NO_3-N	C-7	G-3	G-3	G-3	-	C-7

LEGEND - C - Two continuous composites daily (Night and Day Periods).
G - One grab-sample analyzed daily.
K - Three grab-samples daily at equal intervals (8 hours), combined and analyzed as one.
Number Following Letter - Indicates number of days each week the sample is taken for analysis.

TABLE 6

PHASE III OPERATION - WEEKLY SCHEDULE OF ANALYSES

Analysis	Common Feed Wastewater	Mixed-Liquor						Recycle Sludge		Clarifier Effluent	
		Air System Stages			Oxygen System Stages			Air Clarifier	Oxygen Clarifier	Air System	Oxygen Syster
		No. 1	No. 2	No. 3	No. 1	No. 2	No. 3				
pH	C-7 G-2	G-2	G-2	G-2	G-2	G-2	G-2	-	-	C-7 G-2	C-7 G-2
Turbidity	-	-	-	-	-	-	-	-	-	C-7	C-7
TSS	C-7	K-7 G-2	K-7 G-2	K-7 G-2	K-7 G-2	K-7 G-2	K-7 G-2	K-7 G-7	K-7 G-7	C-7	C-7
VSS	C-7	K-7	K-7	K-7	K-7	K-7	K-7	K-7 G-7	K-7 G-7	-	-
Settleability	-	G-2	G-2	G-2	G-2	G-2	G-2	-	-	-	-
Settling Rate & SVI	-	G-2	G-2	G-2	G-2	G-2	G-2	-	-	-	-
BOD	C-7 G-2	G-2	G-2	G-2	G-2	G-2	G-2	-	-	C-7 G-2	C-7 G-2
COD, TP, TKN, NH_3-N, NO_2-N, and NO_3-N	C-7	G-2	G-2	G-2	G-2	G-2	G-2	-	-	C-7	C-7

LEGEND -	C -	Two continuous composites daily (Night and Day Periods).
	G -	One grab-sample analyzed daily.
	K -	Three grab-samples daily at equal intervals (8 hours), combined and analyzed as one.
Number Following Letter	-	Indicates number of days each week the sample is taken for analysis.

Methods for Individual Measurements

1. Wastewater Feed Flow

Integrating Fischer-Porter magnetic flow meters* were used to monitor raw sewage flow to the air and oxygenation systems. One such meter existed at the plant,measuring total flow to both aeration trains. A second such meter was installed for the contract work down stream of the total flow meter in the leg of the feed line to the air system. Meter totalizers were read twice daily (10:00 a.m. and 11:00 p.m.) to yield "night" and "day" average flow values. Average raw sewage flow to the oxygenation system was obtained as the difference between total wastewater flow and flow to the air train. Raw sewage flow rates to air and oxygen systems were kept in balance as desired by the adjustment of relative liquid levels in each system using effluent mixed-liquor weirs in both sets of aeration tanks. As an example, the flow rate through the oxygenation system could be increased by a slight increase in the height of the air system effluent mixed-liquor overflow weir.

2. Sludge Recycle Flows

Integrating magnetic flow meters in both the air and oxygenation system recycle sludge lines recorded the volume of sludge recycled. By necessity,the sludge return lines for the air and oxygenation treatment systems were completely independent permitting monitoring of each flow independently. Flow integrator readings were taken twice daily as in the case of raw sewage flow. In Phase II of operation,the 4 inch recycle sludge meter in the oxygen system was replaced with a 10 inch meter to permit operation of only one aeration bay with both clarifiers.

3. Waste Activated Sludge Flow

An extra Fischer-Porter magnetic flow meter* was installed in addition to the existing one at the plant so that sludge wasting from the air and oxygenation systems might be independently monitored. Totalized integrator readings were taken once daily to record the volume of sludge wasted.

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* Mention of a proprietary device does not imply endorsement by the Federal Water Quality Administration.

4. Compressed Air Utilized

The compressed air system existing at the plant consisted of two on-line blowers (used alternately) and a spare. The initial two blowers supply air through a single line to air lift pumps and diffusers. Excess air not utilized when half the plant was operated on oxygen was vented to the atmosphere through an orifice flow meter. Revolution counters on each blower were read twice daily (10:00 a.m. and 11:00 p.m.). The average actual flow of compressed air for "night" and "day" periods was calculated from the difference between total displacement at the end and start of each period. Blower discharge pressures and temperatures were recorded and used to convert actual compressed air flow rates to NTP (70°F, 14.7 psia) flow rates. The pressure drop across the excess air vent orifice was recorded as were the pressure and temperature of air on the up-stream side of the orifice. The vented air flow rate was calculated using the orifice calibration constant. Air flow rates to the air lift pumps were similarly estimated. The compressed air utilized in aeration of the activated sludge for a given time period was calculated from:

$$CAU = C - P - V$$

where: CAU = net compressed air utilized

C = air compressed

P = air used for lift pumps

V = vented or wasted excess air

5. Feed Oxygen Flow

A totalizing positive displacement gas flow meter in the feed oxygen line was read twice daily (at 10:00 a.m. and 11:00 p.m.) yielding total flow volumes corresponding to "night" and "day" periods. Averaged feed oxygen pressures and temperatures were taken in each time period and were used to convert feed oxygen flow from actual to NTP (70°F, 14.7 psia) conditions.

6. Waste Oxygen Flow

A totalizing positive displacement gas flow meter in the exhaust oxygen line from the final stage of the system was read twice daily as indicated for feed oxygen. Waste oxygen gas pressures and temperatures were recorded continuously. Using actual cubic feet of oxygen wasted and average temperature and pressure of that gas, the volume was converted to NTP conditions.

7. Gas Recirculation Rates

The rate of oxygen recirculation from each compressor in each stage of the oxygenation system was measured with orifice flow meters. The pressure drop across the orifice, up-stream gas pressure and recirculating gas temperature were measured and recorded continuously in the "night" and "day" time intervals. Oxygen recirculation rates were calculated from:

$$W = K\,(\Delta p \cdot d)^{1/2}$$

where: W = weight rate of discharge
K = orifice calibration constant
Δp = pressure drop across the orifice
d = density of gas up-stream of the orifice

8. Power Required for Gas Recirculation

The power required for oxygen recirculation in each stage of the oxygenation system was calculated from the gas recirculation rate, gas inlet pressure and compressor discharge pressure by assuming single stage adiabatic compression at an efficiency of 70 percent.

9. Oxygen Concentration in Feed, Exhaust, & Stagewise Recirculating Gas

Feed oxygen gas was assumed to be the purity of the temporary liquid supply from which it was vaporized. Oxygen concentrations in the gas space of all oxygenation system stages was continuously monitored using a multipoint Beckman Model F-3 paramagnetic oxygen analyzer* having an accuracy in this range of ± 1/2 percent. Exhaust gas was assumed to have the same composition as that in the gas space of the final stage of the system.

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10. Power Required for Liquid Mixing

Power required for liquid mixing in each stage of the oxygenation system was calculated from wattmeter readings taken several times daily (for each mixer drive motor) using motor efficiency values provided by the manufacturer.

11. Dissolved Oxygen Measurement

The dissolved oxygen concentration at three to five positions (depending upon the number of bays in operation) in the air system mixed-liquor, each operating stage of the oxygenation system mixed-liquor, and in the final settler effluent from each system was monitored continuously. This was accomplished using a multipoint dissolved oxygen monitoring system provided by the Instruments Division of Union Carbide Corporation*. In-place probes mounted at a liquid depth of approximately 18 inches were used in each position. The probes were calibrated in place using a Weston and Stack Model 300E portable dissolved oxygen meter* having a sensing probe, submersible pump assembly, as a secondary standard.

12. Turbidity

This measurement was made for final clarifier effluents routinely and on mixed-liquor (settled) supernatant fractions periodically. The instrument used was a Hach "DR" colormeter-turbidometer*. The scale calibration was in Jackson Turbidity Units (J.T.U.). Standards were prepared as required using the hydrozine sulfate-hexamethyl tetramine standard solution preparation.

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13. BOD

The method employed for BOD measurement was that described in "Standard Methods for the Examination of Water and Wastewater".(11) In all instances, three dilutions per sample were prepared and evaluated. Dissolved oxygen values before and after incubation were determined using a Weston and Stack Model 300E Laboratory dissolved oxygen analyzer* with the Model A-30 BOD Agitator assembly. In each instance, the D.O. analyzer was standardized and calibrated against Winkler titration values for replicate samples. The instrument was highly reliable, as revealed by comparison tests run during the course of the work, and permitted processing of significantly more samples than would have been possible if titration had been done in each instance. All BOD values reported here are from five day BOD determinations only.

14. Suspended Solids and Volatile Suspended Solids

These assays were performed on raw waste, mixed-liquor, recycle sludge and final clarifier effluent routinely using essentially the protocol outlined in "Standard Methods".(11) The only modification was the use of glass fiber filters (H. Reeve Angel & Company, Grade 934AH)* instead of gooch crucibles to collect the residue. The tared filter and residue were dried at 104°C overnight before weighing for total suspended solids. The same tared filter and residue was then fired at 600°C in a muffle furnace, cooled in a dessicator and weighed to permit calculation of the volatile suspended solids by difference. Blank glass fiber filters carried through this procedure indicated a weight loss error of only ± 0.2 mg., an error well within the sensitivity range of the balance used in weighing.

15. Settleability and Settling Rate

This measurement was made using the method given in "Standard Methods"(11). Thirty minute values were used for calculation of SVI (Sludge Volume Index). Five minute interval values were recorded from 0 to 30 minutes to yield initial settling rate data.

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16. Mixed-Liquor Supernatant Analysis

Aliquots of mixed-liquor supernatant were withdrawn after thirty minutes settling in the one liter cylinder to provide samples for the various analyses to be done. For mixed-liquor supernatant (non-filtered) this included COD, BOD, NH_3-N, TKN, TP, NO_2-N, and NO_3-N. These analyses were done to determine relative removal rates through the aeration system stages.

17. COD Determination

The method used for COD analysis was basically that described in "Standard Methods"(11). In cooperation with the analytical services group of the FWQA Robert A.Taft Water Research Center at Cincinnati the following modifications were made to facilitate the processing of large groups of samples. This method was validated by analysis of known COD standards provided by the Robert A.Taft Center Water Research Laboratory.

A 10 ml sample of raw sewage, mixed-liquor supernatant or clarifier effluent is added to a 50 ml screw cap (Teflon cap) test tube. To this sample is added 20 ml of the standard Ag_2SO_4 containing sulfuric acid reagent with mixing. Following this, 10 ml of standardized 0.1N $K_2Cr_2O_7$ solution containing 0.25 gm of $HgSO_4$ is added with mixing. The sample tube is securely capped and placed in an autoclave where it is digested at a pressure of 15 psig and temperature of 120°C for a period of four hours. After removal of the sample from the autoclave and cooling, the sample is quantitatively transferred to a titration flask yielding a final volume with distilled water washing of 200 ml. COD of the sample is then determined by back titration of the $K_2Cr_2O_7$ using 0.1N $Fe(NH_4)_2(SO_4)_2$. The standard ferroin indicator solution is used to determine the titration end point.

18. Total Kjeldahl Nitrogen Determination

This analysis was conducted using a Technicon Autoanalyzer. Initial experimentation, however, indicated the necessity for predigestion of samples containing particulate matters before use of the Technicon system. To accomplish this, 20 ml samples of raw sewage, clarifier effluent, or mixed-liquor supernatant were placed in 50 ml volume screw cap tubes (Teflon cap) and 20 ml of a 33% H_2SO_4 solution was added to each with mixing. The sealed sample tubes were then autoclaved at 15 psig and 120°C for 4 hours. This predigestion step completely solubilized all particulate material assuring the admission of a homogeneous sample to the Technicon Autoanalyzer*.

The predigested sample was then analyzed for total Kjeldahl nitrogen using the Technicon Autoanalysis apparatus using the method described in Technicon Autoanalyzer Methodology, Bulletin N-3c. The only modification of the method as described was the addition of 3.5% $HgCl_2$ solution to the sample prior to passage through the spiral digestion tube. This modification was suggested by the Robert A. Taft Water Research Center of the FWQA as necessary for complete conversion of organic nitrogen to ammonia in addition to the H_2SO_4, SeO_2, and perchloric acid mixture normally used in this digestion step of the Technicon analysis procedure. Accuracy of the method described was verified with TKN standard solution analysis.

19. Ammonia Analysis

Analysis of wastewater samples for ammonia was done using the Technicon Autoanalyzer. The method used was that described in Technicon Bulletin N-3c for TKN analysis except that the digestion step was eliminated and the sample was directly treated.

20. Total Phosphate Analysis

The method used for total phosphate analysis was essentially that described in "Standard Methods"(11), the Aminoaptholsulfamic Acid Method for Orthophosphate. The uniformity of the wastewater sample was assured by digesting it at 15 psig and 120°C in an autoclave for four hours in the presence of sulfuric acid and ammonium persulfate. Subsequent segments of the analytical procedure were carried out using a Technicon Autoanalyzer. This procedure essentially involved conversion of all organically bound phosphate to orthophosphate by heating the dissolved sample at 95°C in the presence of 10N sulfuric acid. The sample was then treated with acid ammonium molybdate to form molybdophosphoric acid which yields an intense blue color complex by the addition of a solution containing 1-amino-2-napthol-4-sulfamic acid, sodium bisulfite and sodium sulfite. The blue color complex is developed by heating at 95°C and is measured spectrophotometrically at 660 mμ.

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* Mention of a proprietary device does not imply endorsement by the Federal Water Quality Administration.

21. Nitrite and Nitrate Analysis

These analysis were carried out through an automated procedure using the Technicon Autoanalyzer. The automated method is essentially the Diazotization Method described in the FWQA Training Manual for Laboratory Analysis in treatment plant operation. For total nitrite analysis, the sample is treated directly with the sulfanilic acid and α-napthylamine to form an intense red azo dye. The color intensity was measured spectrophotometrically at 520 mμ.

Analysis for nitrate in the sample employs the foregoing procedure after the nitrate is reduced to nitrite by treatment with hydrazine sulfate in an alkaline medium heated to 53°C. Nitrate nitrogen is measured as the difference between values obtained for total nitrite alone and reduced nitrate (nitrite plus total nitrate nitrogen).

Data Reduction

In day-to-day operation throughout the course of the work only raw data from analysis and monitoring was collected at the work site. In weekly batches this data was computationally reduced at Tonawanda using an IBM 360/50 computer* and a program specifically written to analyze and reduce the raw data to meaningful form. Technicon autoanalysis results were dealt with in a similar manner. The product of this approach to data reduction was a weekly average and daily listing of all system operating conditions and performance results. The most beneficial result of this means of data reduction was that it permitted significantly more observations and analyses to be made in the work since technicians were relieved of devoting time to calculating and cataloging results. The quantity of data obtained for each day of operation is so extensive that they will not be listed on a daily basis in this report. Weekly averages only will be listed in the Results section.

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* Mention of a proprietary device does not imply endorsement by the Federal Water Quality Administration.

RESULTS

Performance Comparison of Air & Oxygenation Systems at Low Treatment Rate & Low Mixed-Liquor Solids Concentrations - Phase No. 1 Operation

The parallel operation of the air and oxygenation systems at Batavia in the initial phase of evaluation has been shown schematically in Figure 4. Following the completion of construction work at the site, operation of the temporarily installed oxygen aeration system was initiated on April 1, 1969. This system was operated separately, but in parallel with the air aerated conventional activated sludge treatment unit existing at the plant. The two units were completely separate sharing equally the raw wastewater flow into the plant.

At design flow (2.5 MGD, or 1.25 MGD per aeration train) the two aeration units should have operated at an aeration detention time (based on raw sewage flow) of approximately 6 hours in each case (325,200 gallon aeration tank volume per aeration train). The Spring of 1969, however, was one of unusually heavy rainfall in Western New York such that raw sewage inflow to the plant often was 2-3 times design for sustained periods of time. This condition did not cause difficulty in the aerators of either system, but did introduce substantial problems of hydraulically overloading the final clarifiers. Sustained overflow rates in April of 2400 gal/ft^2/day or greater were not unusual. Further, the maximum pumping capacity of the air lift sludge return pumps was such that an average recycle sludge flow/raw sewage inflow ratio of greater than 0.13 was not usually possible during this period. The resulting frequent flushing of solid from the final clarifiers severely limited the acquisition of meaningful performance data during April. For this reason, these limited results are not reported here.

To improve clarifier operation, new air lift sludge return pumps were installed in the first week of May which effectively doubled the maximum recycle sludge capacity. This action along with a decrease in plant hydraulic loadings during May permitted the consistent acquisition of meaningful comparative test data in the second week of that month. The oxygenation system had been operating for approximately six weeks without problems at that time.

Following installation of higher sludge recycle capacity in May, no further significant difficulties were encountered during the initial phase of operation (through June 29, 1969). All process monitoring instrumentation for both the air and oxygen systems functioned reliably with only one exception, the recycle sludge magnetic flow meters. Partial clogging of these meters with rags and other debris yielded erroneous readings on occasion. Also the inability to properly clean the sensing elements in these meters caused some variations in sensitivity. Both of these difficulties probably stem from the lack of primary clarification of raw sewage entering the aeration systems. This relatively minor metering problem did not detract significantly from the value of the data presented here.

The oxygenation system equipment, described earlier, functioned in a highly reliable manner during this phase of operation. Following start-up of the system and adjustment of feed oxygen and dissolved oxygen controls, no further changes were necessary to operate at the predetermined conditions. These conditions were basically treatment of one-half the influent wastewater in a six stage oxygenation unit at a 3000-4000 mg/l mixed-liquor suspended solids (MLSS) concentration and a mixed-liquor dissolved oxygen concentration (D.O.) of 8-10 mg/l with an overall oxygen utilization efficiency of greater than 90%.

The air aeration system was operated conventionally with the exception that the addition of 50% more diffusers in the tanks permitted higher aeration rates than normally practiced at this plant. In general, the air aeration rate was controlled in the air system to provide a mixed-liquor D.O. of approximately 2.0 mg/l throughout the aeration tanks. The use of manual valves on each diffuser header permitted tapering of the aeration rate along the length of the tank. In this manner a mean aeration rate of 3.0 ft^3/gal. of sewage treated was achieved. The air input rate at the inlet end of the aeration tank was approximately double that of the outlet end. This yielded an almost even D.O. concentration throughout the aeration tanks.

As noted from the operating conditions shown in Tables 7 and 8, this initial mode of operation was a relatively conservative one compared to existing practices at other facilities. These data reported here are weekly averages of daily values obtained continuously on a 24 hour basis. The combination of these seven weeks of continuous data, after six weeks for start-up and development of characteristic biomass in each system, provided the basis for a valid performance evaluation.

As shown in Table 7, feed wastewater flow rates were maintained at approximately equivalent values averaging 1.97 MGD for the air system and 1.91 MGD for the oxygenation unit for the seven week period. Approximate control of flow through the aeration systems was accomplished by adjustment of mixed-liquor effluent weirs installed for this purpose. Clarifier operation was not unusual with recycle sludge to feed sewage flow ratios in the conventional range. Recycle sludge concentrations averaging 14,964 mg/l and 18,620 mg/l were observed for the air and oxygenation units,respectively,during this operation period. Mixed-liquor D.O. concentrations were maintained at an average of 1.5 and 8.7 mg/l for the air and oxygen systems,respectively. In the case of the oxygen system, D.O. control was such that a range of 8 ± 2 mg/l was maintained in the mixed-liquor throughout the study.

As illustrated in Table 9,a mean aeration rate of 2.89 ft^3/gal. of sewage treated was maintained in the air system using a net average air blower power input of 141 HP throughout the seven weeks. In the oxygen system, pure oxygen was fed to the unit at an average rate of 18.3 cfm

TABLE 7

PHASE I OPERATION - AIR AND OXYGEN SYSTEMS PERFORMANCE COMPARISON

Weekly Average of Daily Values

PARAMETER	PERIOD	5/12/69		5/19/69		5/26/69		6/2/69		6/9/69		6/16/69		6/23/69	
		AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN
Wastewater Feed Rate in MGD	Night	2.03	1.57	2.25	2.35	1.79	1.82	1.76	1.80	1.57	1.62	1.52	1.54	1.64	1.18
	Day	2.20	1.88	2.39	2.59	2.08	2.17	2.13	2.04	2.03	2.20	1.92	1.99	2.06	1.75
	24 Hour Mean	2.12	1.74	2.33	2.48	1.95	2.01	1.96	1.93	1.82	1.93	1.74	1.79	1.87	1.49
Sludge Recycle Rate in MGD	Night	0.31	0.61	0.27	0.49	0.26	0.32	0.28	0.43	0.21	0.30		0.29		0.44
	Day	0.26	0.60	0.27	0.50	0.25	0.31	0.21	0.43	0.22	0.47		0.43		0.51
	24 Hour Mean	0.28	0.61	0.27	0.49	0.25	0.31	0.24	0.43	0.22	0.39		0.36		0.48
Sludge Recycle to Raw Waste Feed Rate Ratio	Night	0.15	0.42	0.12	0.20	0.14	0.18	0.16	0.24	0.13	0.19		0.19		0.38
	Day	0.12	0.34	0.11	0.19	0.12	0.14	0.10	0.21	0.11	0.21		0.22		0.29
	24 Hour Mean	0.14	0.38	0.12	0.20	0.13	0.16	0.13	0.22	0.12	0.20		0.21		0.33
Wastewater Temp. °F	Daily Average	53	53	53	53	54	54	61	61	63	63	65	65	67	67
Mixed Liquor DO as Average mg/l	Night	3.5	10.5	2.5	11.9	1.5	10.8	0.1	12.3	2.3	8.6	1.3	9.3	2.7	8.0
	Day	0.4	9.4	0.8	9.9	0.2	6.5	1.2	8.9	0.1	5.1	1.0	5.0	2.6	5.5
	24 Hour Mean	1.9	10.1	1.6	10.8	0.8	8.8	0.8	10.5	1.2	6.6	1.2	7.0	2.7	6.9
Clarifier Effluent DO in mg/l	Night	1.7	3.9	0.8	4.0	1.1	7.2	0.1	8.7	0.0	5.3	0.0	5.4	0.0	7.0
	Day	0.4	2.2	0.4	0.6	0.9	2.6	0.2	5.1	0.0	2.3	0.1	2.3	0.0	2.5
	24 Hour Mean	1.0	2.9	0.6	2.1	1.1	4.8	0.1	6.6	0.0	3.5	0.1	3.8	0.0	4.1
Aeration Detention Time in hours (Wastewater & Recycle)	Night	3.3	3.6	3.1	2.8	3.8	3.7	3.9	3.5	4.4	4.1		4.3		4.9
	Day	3.2	3.1	3.0	2.5	3.4	3.2	3.4	3.2	3.5	2.9		3.2		3.5
	24 Hour Mean	3.3	3.3	3.0	2.7	3.6	3.4	3.6	3.3	3.9	3.5		3.7		4.1
Nominal Aeration Detention Time in hours (Wastewater Only)	Night	3.8	5.0	3.5	3.3	4.4	4.3	4.4	4.3	5.0	4.8	5.1	5.1	4.8	6.6
	Day	3.5	4.2	3.3	3.0	3.8	3.6	3.7	3.8	3.8	3.5	4.0	3.9	3.8	4.5
	24 Hour Mean	3.7	4.5	3.3	3.1	4.0	3.9	4.0	4.0	4.3	4.0	4.5	4.4	4.2	5.2
Raw Wastewater pH	Daily Average		7.5		7.9		7.7		8.5		7.9		8.1		
Mixed Liquor pH	Inlet Stg. 1	7.6	7.4	7.7	7.5	7.7	7.6	6.8	6.6	7.4	7.1	7.5	7.2	7.9	7.5
	Middle Stg. 4		7.2		7.3		7.3		6.4		6.8		6.9	7.8	7.2
	Outlet Stg. 6	7.6	7.1	7.7	7.3	7.7	7.2	6.8	6.3	7.2	6.7	7.4	6.8	7.8	7.1
Clarifier Effluent pH	Daily Average					8.0	7.5	7.0				7.7	7.7		

TABLE 8

PHASE I OPERATION - AIR AND OXYGEN SYSTEMS PERFORMANCE COMPARISON

PARAMETER	PERIOD	5/12/69		5/19/69		5/26/69		6/2/69		6/9/69		6/16/69		6/23/69	
		Weekly Average of Daily Values													
		AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN
VSS of Feed Wastewater (mg/1)	Night Day 24 Hour Mean	131 178 173	131 178 173	68 154 118	68 154 118	95 267 198	95 267 198	86 207 163	86 207 163	79 167 130	79 167 130	103 179 149	103 179 149	184 130 136	184 130 136
Mixed Liquor TSS in mg/1	24 Hour Mean	2670	3120	2490	2940	2570	3210	2480	3150	2030	3500	2580	3270	2280	3250
Mixed Liquor VSS in mg/1	24 Hour Mean	1740	2200	1690	1900	1730	2220	1740	2178	1720	2410	1720	2540	1840	2400
Recycle Sludge TSS in mg/1	24 Hour Mean	11740	14710	13740	21000	15240	18420	15190	19850	14810	18600	17110	19220	16930	18540
Recycle Sludge VSS in mg/1	24 Hour Mean	7500	8520	8680	12860	10160	12000	9690	12260	9901	11750	11310	12660	10710	12280
Recycle Sludge VSS/TSS Ratio	24 Hour Mean	0.64	0.59	0.63	0.61	0.65	0.66	0.64	0.62	0.67	0.63	0.66	0.66	0.63	0.65
Sludge Volume Index	24 Hour Mean	87	60	84	54	76	55	62	59	60	65	74	70	91	87
Initial Sludge Settling Rate in ft/hr	24 Hour Mean	7.0	6.4	8.0	9.4	7.8	7.1	9.2	8.1	9.0	6.7	6.8	6.3	6.0	6.2
lbs. Dry Solids Solids Wasted	Each 24 Hours	4350	2110	2060	1100	3290	1860	2690	1780	2060	2470	3790	2330	3800	980
lbs. VSS Wasted per lb. BOD Removed	24 Hour Mean	1.10	0.59	0.74	0.35	0.62	0.34	0.83	0.52	0.73	0.73	1.32	0.77	0.87	0.29
Food/Biomass as lb. BOD per lb. MLVSS	Night Day 24 Hour Mean	0.484 0.721 0.668	0.263 0.472 0.420	0.213 0.627 0.500	0.197 0.602 0.408	0.478 0.843 0.801	0.378 0.684 0.642	0.289 0.701 0.509	0.236 0.536 0.401	0.264 0.647 0.473	0.194 0.468 0.356	0.266 0.534 0.418	0.193 0.402 0.312	0.422 0.726 0.593	0.232 0.472 0.362
Volumetric Organic Loading as lbs. BOD per 1000 cf of Mixed Liquor	Night Day 24 Hour Mean	46.1 74.4 60.6	32.1 63.4 47.0	22.9 67.9 47.5	24.4 76.2 52.2	53.3 91.4 82.0	51.3 100.2 90.7	31.2 75.8 55.4	31.9 73.3 54.3	28.5 69.9 50.9	29.4 74.7 53.9	31.3 60.0 48.5	32.3 63.7 51.9	47.9 85.0 68.0	34.4 72.3 55.0
Effluent Turbidity as J.T.U.	Night Day 24 Hour Mean			2.3 2.5 2.4	2.2 2.6 2.8	2.8 3.9 3.4	4.4 4.9 4.8	9.4 7.7 8.4	4.1 6.2 4.5	8.0 8.2 8.1	4.4 8.2 6.7	7.9 7.8 7.8	8.0 13.5 11.3	3.5 4.3 4.0	4.3 3.1 3.6

TABLE 9

PHASE I OPERATION - AIR AND OXYGEN SYSTEMS PERFORMANCE COMPARISON

Weekly Average of Daily Values

PARAMETER	PERIOD	5/12/69		5/19/69		5/26/69		6/2/69		6/9/69		6/16/69		6/23/69	
		AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN
Ft^3 Air Utilized per Gallon Sewage Treated (NTP)	Night	2.76		2.61		3.18		3.04		3.93		3.84		3.93	
	Day	2.37		2.31		2.46		2.42		2.93		2.84		2.76	
	24 Hour Mean	2.54		2.44		2.76		2.67		3.33		3.24		3.23	
Air Blower Power Utilized as HP	Night	138.6		145.7		141.2		134.6		156.0		147.8		162.9	
	Day	128.8		136.5		126.7		129.8		150.7		137.4		143.6	
	24 Hour Mean	133.3		140.7		133.3		132.0		153.1		142.2		152.5	
Total Feed Oxygen Flow Rate, CFM (NTP)	Night		15.8		13.6		16.2		17.1		18.1		18.9		19.5
	Day		16.6		16.6		18.8		18.5		21.3		21.6		22.3
	24 Hour Mean		16.2		15.2		17.6		17.9		19.9		20.3		21.0
Average % Oxygen Composition in Gas Space of Each Oxygen System Stage	24 Hour Mean No. 1		92		90		91		92		94		91		91
	No. 2		84		81		83		86		87		85		85
	No. 3		80		77		80		82		83		81		81
	No. 4		73		68		72		76		78		75		76
	No. 5		64		57		62		65		66		65		63
	No. 6		49		42		42		46		48		47		45
Exhaust Gas Rate from Oxygen System 6th Stage cfm(NTP)	Night		2.53		2.13		1.46		1.40		1.48		1.58		1.61
	Day		2.55		2.18		1.57		1.40		1.52		1.46		1.49
	24 Hour Mean		2.40		2.16		1.52		1.40		1.50		1.52		1.54
Overall % of Feed Oxygen Utilized	Night		93.3		93.2		96.2		96.2		95.8		96.1		96.3
	Day		92.1		94.5		96.5		96.5		96.6		96.8		96.9
	24 Hour Mean		92.6		94.0		96.4		96.4		96.2		96.5		96.7
lbs. Oxygen Utilized per lb. of BOD Consumed	Night		1.20		1.79		0.91		1.63		1.84		1.76		1.61
	Day		0.68		0.65		0.55		0.74		0.81		0.95		0.88
	24 Hour Mean		0.85		0.90		0.56		0.96		1.05		1.18		1.09
Oxygen Absorbed as lbs/hr	Night		73.4		63.1		77.6		81.9		85.6		90.3		93.3
	Day		75.6		77.9		90.2		88.9		102.4		103.6		107.3
	24 Hour Mean		74.6		71.1		84.4		85.7		94.3		97.5		100.9

and gas was exhausted from the system at an average rate of 1.74 cfm. This exhaust gas contained an average of 46% oxygen. On this basis, the overall average oxygen utilization efficiency for the oxygenation system during the seven week period was 95.5%. As shown in Table 10, power required to accomplish liquid mixing and gas recirculation in the oxygen system averaged 26.0 and 2.6 HP, respectively. Individual power consumption values for each aeration stage are listed on a weekly average basis. The total average power input for the oxygenation system was 28.6 HP for the operational period shown. On a unit tank volume basis, the seven week average of those values listed in Table 10 indicates a power input for the oxygenation unit of 0.088 HP/1000 gal. of mixed-liquor.

From the standpoint of treatment effectiveness, the results shown in Table 11 list the influent waste strength in terms of BOD, COD, and total suspended solids (TSS). Influent volatile suspended solids concentrations are noted in Table 8. Also shown in Table 11 are the concentrations of BOD, COD, and TSS in the final clarifier effluent after treatment. The BOD and COD concentrations of the clarifier effluent were determined by analysis of whole, not filtered, samples. Listed appropriately are weekly average values for BOD, COD, and TSS removal throughout the seven week period. As in most of the data presented here, night, day and 24-hour average (flow weighted) values are listed in each instance. The time period for night samples was pointed out earlier as 11:00 p.m. to 10:00 a.m. the following morning with day samples including continuous composites taken from 10:00 a.m. to 11:00 p.m. This latter period corresponds to the highest organic and hydraulic loading period of the day.

As illustrated by examination of the data in Table 11, the waste strength with respect to BOD was lower during this seven week period than the normal yearly average at the plant of 200 mg/l. This, no doubt, was due to the dilution effect of ground water infiltrating the system. The seven week average BOD of the influent wastewater was 159 mg/l on a 24-hour basis. The clarifier effluents for the air and oxygenation

TABLE 10

PHASE I OPERATION - AIR AND OXYGEN SYSTEMS PERFORMANCE COMPARISON

		Weekly Average of Daily Values													
PARAMETER	PERIOD	5/12/69		5/19/69		5/26/69		6/2/69		6/9/69		6/16/69		6/23/69	
		AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN
Power Required for Liquid Mixing in Oxygen System (per Stage as HP)	24 Hour Mean No. 1		7.3		7.1		6.6		7.2		7.0		7.1		7.2
	No. 2		4.3		4.6		4.5		4.9		4.8		4.6		4.9
	No. 3		4.3		4.2		4.1		4.2		4.1		3.9		4.2
	No. 4		3.9		3.8		3.8		3.9		3.8		3.7		3.8
	No. 5		3.1		3.1		3.0		3.1		3.1		3.1		3.1
	No. 6		3.2		3.1		3.0		3.2		3.2		3.0		3.2
Power Required for Gas Recirculation in Oxygen System (per Stage as HP)	24 Hour Mean No. 1		0.5		0.7		0.7		0.7		0.8		0.6		0.8
	No. 2		0.5		0.6		0.5		0.6		0.5		0.6		0.6
	No. 3		0.4		0.3		0.3		0.4		0.3		0.4		0.4
	No. 4		0.3		0.3		0.3		0.3		0.3		0.3		0.3
	No. 5		0.3		0.3		0.3		0.4		0.3		0.3		0.4
	No. 6		0.3		0.4		0.4		0.4		0.3		0.4		0.5
Overall Power Required for Liquid Mixing (as HP)	Night		26.6		25.9		25.3		26.3		25.5		25.3		26.1
	Day		26.5		25.9		24.8		26.6		26.4		25.6		26.6
	24 Hour Mean		26.6		25.9		25.0		26.5		26.0		25.5		26.4
Overall Power Required for Gas Recirculation (as HP)	Night		2.2		2.6		2.5		2.9		2.4		2.7		3.0
	Day		2.2		2.6		2.3		2.9		2.4		2.5		2.9
	24 Hour Mean		2.2		2.6		2.4		2.9		2.4		2.5		3.0
Overall Power Utilized as HP per 1000 ft^3 of Mixed Liquor	Night		0.089		0.088		0.085		0.090		0.086		0.086		0.089
	Day		0.088		0.088		0.083		0.091		0.087		0.086		0.091
	24 Hour Mean		0.089		0.088		0.084		0.090		0.087		0.086		0.090
Clarifier Overflow Rates as gal/ft^2/day	Night	1616	1250	1791	1870	1425	1448	1401	1433	1250	1289	1210	1226	1305	939
	Day	1751	1496	1902	2061	1655	1727	1695	1624	1616	1751	1528	1584	1640	1393
	24 Hour Mean	1687	1385	1854	1978	1552	1600	1560	1536	1448	1536	1385	1425	1488	1186

TABLE 11

PHASE I OPERATION - AIR AND OXYGEN SYSTEMS PERFORMANCE COMPARISON

		Weekly Average of Daily Values													
		5/12/69		5/19/69		5/26/69		6/2/69		6/9/69		6/16/69		6/23/69	
PARAMETER	PERIOD	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN
Feed Wastwater BOD (mg/l)	Night	120	120	52	52	150	150	93	93	94	94	105	105	154	154
	Day	180	180	144	144	228	228	186	186	178	178	167	167	211	211
	24 Hour Mean	152	152	102	102	231	231	147	147	145	145	144	144	190	190
Clarifier Effluent BOD (mg/l)	Night	10	9	10	9	18	16	16	13	18	11	13	10	16	12
	Day	11	7	13	10	22	11	23	17	23	12	14	9	13	10
	24 Hour Mean	10	7	12	11	19	14	19	14	21	12	14	11	14	11
% BOD Removed	Night	91	93	81	83	88	89	82	86	81	88	88	91	90	92
	Day	94	96	91	93	90	95	88	91	87	93	91	95	94	95
	24 Hour Mean	93	95	88	90	92	94	87	90	86	92	90	92	93	94
Feed Wastewater COD (mg/l)	Night	312	312	178	178	262	262	180	180	214	214	217	217	342	342
	Day	560	560	376	376	465	465	398	398	362	362	292	292	419	419
	24 Hour Mean	461	461	289	289	389	389	307	307	304	304	330	330	386	386
Clarifier Effluent COD (mg/l)	Night	189	193	53	47	50	59	72	48	66	62	66	54	90	74
	Day	196	197	32	44	72	60	100	47	74	29	63	45	60	68
	24 Hour Mean	183	197	41	47	63	60	89	45	72	41	62	50	78	74
% COD Removed	Night	40	38	70	74	81	77	60	73	69	71	69	75	74	78
	Day	65	65	92	88	84	87	75	88	80	92	84	89	86	84
	24 Hour Mean	58	57	86	84	84	85	71	85	76	86	81	85	80	81
Feed Wastewater TSS (mg/l)	Night	132	132	101	101	149	149	110	110	98	98	137	137	261	261
	Day	317	317	244	244	386	386	333	333	210	210	240	240	184	184
	24 Hour Mean	267	267	185	185	286	286	239	239	167	167	210	210	192	192
Clarifier Effluent TSS (mg/l)	Night	9	13	12	8	10	12	21	8	20	8	15	11	13	11
	Day	12	10	10	12	13	8	22	18	26	8	21	6	13	9
	24 Hour Mean	10	5	11	10	12	10	22	14	24	8	19	8	12	10
% TSS Removed	Night	93	90	89	92	93	92	81	93	79	91	89	92	95	96
	Day	96	97	96	95	97	98	93	95	88	96	91	97	93	95
	24 Hour Mean	95	96	94	95	96	96	91	94	86	95	91	96	94	95

systems contained 16 and 11 mg/l BOD, respectively, yielding an average overall BOD removal by the air system of approximately 90% and a removal of 92% by the oxygenation system. The COD concentration of influent wastewater averaged 352 mg/l for the seven week test period on a 24-hour per day basis. As illustrated in Table 11, daytime COD concentrations as high as 560 mg/l on a weekly average basis were observed. Seven week average COD concentrations in the effluent of the air and oxygenation systems of 84 and 73 mg/l, respectively, were observed, indicating an average COD removal efficiency for the test period of 80% for the oxygenation system and 76% for the air aeration unit. TSS removals by both the oxygenation system and air aerated activated sludge unit were quite good as may be observed in Table 11. From a seven week average TSS concentration of 221 mg/l in the influent wastewater, 93% was removed in the air aeration system and 96% in the oxygenation system.

Table 12 indicates weekly average values for phosphorous removal by the air and oxygenation systems. Removal of 44% for the air system and 39% for the oxygenation system were observed on an overall seven week average basis. These values were based on total phosphate analysis rather than ortho-phosphate alone.

Listed in Tables 12 and 13 are weekly average results of analysis for NH_3-N, TKN, NO_2-N and NO_3-N in feed wastewater to and clarifier effluent from both the air and oxygenation treatment systems. In viewing these results, it should be noted that in the instance of clarifier effluents, whole composite samples rather than filtered samples were analyzed. On an overall seven week basis, NH_3-N removal by the oxygenation system was

TABLE 12

PHASE I OPERATION - AIR AND OXYGEN SYSTEMS PERFORMANCE COMPARISON

PARAMETER	PERIOD	Weekly Average of Daily Values													
		5/12/69		5/19/69		5/26/69		6/2/69		6/9/69		6/16/69		6/23/69	
		AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN
Total Phosphate in Feed Wastewater as mg/l of P	Night	9	9	5	5	8	8	6	6	6	6	8	8	10	10
	Day	15	15	12	12	17	17	15	15	12	12	17	17	12	12
	24 Hour Mean	12	12	9	9	13	13	12	12	10	10	14	14	10	10
Total Phosphate in Clarifier Effluent as mg/l of P	Night	5	6	4	5	5	6	5	6	7	8	8	9	6	7
	Day	7	7	5	6	7	7	6	6	7	8	8	8	7	8
	24 Hour Mean	6	7	5	5	6	7	6	6	7	8	8	9	7	7
Total % Phosphate Removed	Night	38	30	24	13	36	22	16	13	-10	-27	-1	-19	40	31
	Day	52	51	55	49	61	57	58	59	41	37	55	51	38	36
	24 Hour Mean	49	45	47	42	56	49	48	48	29	21	45	39	32	30
Ammonia Nitrogen in Feed Wastewater as mg/l N	Night	14	14	10	10	26	26	24	24	16	16	19	19	25	25
	Day	17	17	15	15	22	22	23	23	22	22	25	25	21	21
	24 Hour Mean	16	16	13	13	23	23	23	23	20	20	24	24	23	23
Ammonia in Clarifier Effluent as mg/l N	Night	17	17	9	10	16	12	18	18	18	18	23	21	16	13
	Day	13	14	12	11	19	5	21	19	19	18	19	16	20	19
	24 Hour Mean	13	14	10	10	18	8	20	19	18	18	21	19	20	18
TKN of Feed Wastewater as mg/l N	Night	27	27	19	19	44	44	28	28	25	25	27	27	36	36
	Day	35	35	32	32	48	48	38	38	35	35	38	38	34	34
	24 Hour Mean	31	31	26	26	45	45	34	34	31	31	35	35	35	35
TKN of Clarifier Effluent as mg/l N	Night	21	22	16	13	21	20	23	19	18	18	26	23	17	14
	Day	19	18	19	16	26	20	25	22	19	19	24	18	23	20
	24 Hour Mean	19	19	17	15	24	20	24	20	19	18	26	21	22	20
NO_2 Nitrogen in Feed Wastewater as mg/l N	Night	0.2	0.2	0.4	0.4	0.3	0.3	0.4	0.4	0.1	0.1	0.1	0.1	0.1	0.1
	Day	0.2	0.2	0.3	0.3	0.0	0.0	0.2	0.2	0.0	0.0	0.2	0.2	0.0	0.0
	24 Hour Mean	0.2	0.2	0.3	0.3	0.1	0.1	0.3	0.3	0.1	0.1	0.1	0.1	0.0	0.0
NO_2 Nitrogen in Clarifier Effluent as mg/l N	Night	0.9	0.5	0.9	0.8	0.4	1.1	0.2	1.0	0.1	0.9	0.3	1.1	0.1	1.1
	Day	0.6	0.6	1.0	0.9	0.4	0.9	0.2	0.9	0.1	0.7	0.1	0.8	0.2	1.2
	24 Hour Mean	0.7	0.5	1.0	0.8	0.4	1.0	0.2	0.9	0.1	0.8	0.2	0.9	0.1	1.2

TABLE 13

PHASE I OPERATION - AIR AND OXYGEN SYSTEMS PERFORMANCE COMPARISON

Weekly Average of Daily Values

PARAMETER	PERIOD	5/12/69		5/19/69		5/26/69		6/2/69		6/9/69		6/16/69		6/23/69	
		AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN
NO$_3$ Nitrogen in Feed Wastewater as mg/l N	Night	0.8	0.8	1.3	1.3	0.5	0.5	0.8	0.8	0.1	0.1	0.1	0.1	0.3	0.3
	Day	0.4	0.4	0.5	0.5	0.1	0.1	0.0	0.0	0.1	0.1	0.1	0.1	0.1	0.1
	24 Hour Mean	0.6	0.6	0.8	0.8	0.2	0.2	0.3	0.3	0.1	0.1	0.1	0.1	0.2	0.2
NO$_3$ Nitrogen in Clarifier Effluent as mg/l N	Night	0.8	1.2	0.9	1.9	0.7	1.5	1.1	2.2	0.9	1.4	1.0	2.1	0.1	3.4
	Day	0.4	1.0	0.6	1.7	0.6	0.9	0.7	1.7	0.8	1.0	0.7	1.3	0.2	2.3
	24 Hour Mean	0.5	1.1	0.7	1.8	0.7	1.2	0.8	2.0	0.8	1.1	0.9	1.5	0.1	2.7
Total Nitrogen in Feed Wastewater as mg/l N	Night	28	28	21	21	45	45	29	29	26	26	27	27	36	36
	Day	35	35	33	33	48	48	39	39	35	35	39	39	35	35
	24 Hour Mean	32	32	27	27	46	46	35	35	31	31	36	36	36	36
Total Nitrogen in Clarifier Effluent as mg/l N	Night	23	24	17	16	22	23	24	22	19	20	27	26	17	19
	Day	20	20	20	18	27	22	26	24	20	20	25	21	23	23
	24 Hour Mean	20	21	19	17	25	22	25	23	20	20	27	23	22	23
% Total Nitrogen Removed	Night	18	14	17	24	50	49	18	26	26	22	-1	2	53	49
	Day	45	43	37	44	44	55	33	38	42	41	36	47	34	33
	24 Hour Mean	38	35	31	36	46	52	28	33	37	35	25	34	37	34
NO$_3$-N/Total N in Feed Wastewater	Night	0.03	0.03	0.06	0.06	0.01	0.01	0.03	0.03	0.00	0.00	0.01	0.01	0.01	0.01
	Day	0.02	0.02	0.02	0.02	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
	24 Hour Mean	0.02	0.02	0.04	0.04	0.01	0.01	0.01	0.01	0.00	0.00	0.00	0.00	0.00	0.00
NO$_3$-N/Total N in Clarifier Effluent	Night	0.04	0.06	0.06	0.14	0.05	0.08	0.04	0.11	0.05	0.07	0.04	0.09	0.00	0.21
	Day	0.02	0.05	0.04	0.09	0.03	0.04	0.03	0.07	0.04	0.05	0.03	0.06	0.01	0.12
	24 Hour Mean	0.03	0.06	0.05	0.11	0.04	0.06	0.03	0.09	0.04	0.06	0.03	0.07	0.00	0.14
Lbs. Total Nitrogen Removed per Lb. Dry Solids Wasted	24 Hour Mean	0.049	0.075	0.075	0.188	0.104	0.216	0.060	0.108	0.081	0.072	0.034	0.083	0.057	0.165

higher than that of the air aerated unit, a concentration of 15 mg/l in the clarifier effluent of the former system and 17 mg/l in the latter. This same comparative relationship existed for TKN, 22 mg/l in air system effluent and 19 mg/l in oxygenation system effluent. It is clear that virtually all TKN of the clarifier effluents in both cases is constituted by NH_3-N rather than organically bound nitrogen. Examination of the nitrite and nitrate nitrogen content of feed wastewater and clarifier effluents reflect the magnitude of difference in NH_3-N and TKN removal between the two systems. The oxygenation system accomplished some nitrification, 0.8 mg/l NO_2-N and 1.6 mg/l NO_3-N in the effluent, while the air aeration system effluent contained 0.4 mg/l NO_2-N and 0.6 mg/l NO_3-N. Overall nitrogen removals by both systems are listed in Table 13 on a weekly average basis. Also listed in each instance is the NO_3-N/Total N ratio of feed wastewater and clarifier effluent and the quantity of nitrogen removed in each system as a function of the quantity of dry solids wasted.

Characteristics of the biomass from the air and oxygenation systems are indicated in Table 8. On an average basis for the seven week period, the MLVSS concentration of the oxygenation system was 2,266 mg/l while that of the air system was 1,740 mg/l. This biomass was subjected to volumetric organic loadings on a 24-hour average of 57.9 and 60.0 lbs. of BOD/1000 ft^3 of mixed-liquor for the oxygenation and air systems, respectively. The corresponding food to biomass daily weight ratios are also listed on a weekly average basis. Day to night variations in food/biomass ratio vary by a factor of two on the overall average and by as much as a factor of six in a given week for both systems. The seven week average SVI of the air and oxygenation systems were 76 and 64, respectively. The settling rate of the two activated sludges were nearly identical on an overall average basis, even though the higher MLSS concentration in the oxygenation system should have yielded a relatively lower hindered settling velocity than that of the lower MLSS concentration in the air system. Settling rates were determined using the standard one liter graduated cylinder.

Table 8 also lists the daily amount of dry solids wasted from each treatment system on a weekly average basis. Remarkably, the air system produced a daily average of 3,148 lbs. of solids for the seven week period while a daily average of 1804 lbs. were wasted from the oxygenation system. As indicated from the recycle sludge VSS/TSS ratio, the wasted sludge from both systems is approximately 65% volatile solids. Since both aeration systems received approximately the same quantity of common influent sewage, hence the same amount of influent TSS, VSS, BOD, and COD, it is clear that a characteristic of the oxygenation treatment system is the production of significantly less waste activated sludge than the air aeration system operated in parallel with it. This is further shown in Table 8 wherein comparative values are listed for lbs. of excess VSS wasted from each system per lb. of BOD removed (an average of 0.856 lbs. VSS/lb. BOD in the air system and 0.482 lbs. VSS/lb. BOD in the oxygenation system). In this context, it is useful to note the lbs. of oxygen utilized in the removal of one lb. of BOD in Table 9. This is key to the subsequent discussion of how significant reductions in waste sludge production are possible with oxygen aeration.

During the final week of parallel operation of the air and oxygenation treatment systems in June, a brief period of intensive analysis of the mixed-liquor in each system was carried out. The results of these analyses are listed in Tables 14, 15, 16, 17, and 18. This work included a cursory enumeration of higher biological forms, rotifers and protozoa present in each system. As indicated, grab-samples (rather than composites) were evaluated. All samples were taken at the times indicated.

Included in the Appendix of this report is Table I, analyses of mixed-liquor supernatant from each system made during six weeks of the Phase I study. In this instance, the values result from analysis of mixed-liquor supernatant decanted from the one liter graduated cylinder after the thirty minute settling test.

TABLE 14

INTENSIVE MIXED-LIQUOR EVALUATION

9:00 A.M. - June 24, 1969

Sample[1]	BOD[2]	TSS	VSS	TS	TVS	Turbidity[3]	Settling[4] Velocity	SVI	pH	COD	P	TKN	NH_3-N	NO_2-N	NO_3-N	Free Swim. Cili / Stalked Cilia
Common Feed Waste	130	175	118	-	-	62.0	-	-	8.76	282	6.7	43.0	29.0	0.3	0.1	
Air Train																
Inlet	44	3080	1858	3732	2112	8.4	6.1	87	8.00	105	5.4	26.9	21.8	0.0	0.0	$\frac{290/ml}{60/ml} = 4.8$
Middle	20	3250	1982	3948	2190	7.2	6.0	86	7.87	124	5.4	22.8	18.0	0.0	0.0	
Exit	19	3126	1860	3918	2222	7.3	6.8	97	7.82	141	5.1	19.3	17.5	0.0	0.0	
Oxygen Train																
Stage 1	47	2960	1958	3550	2028	16.0	7.5	85	7.65	111	6.6	34.4	23.1	0.5	0.7	$\frac{480/ml}{70/ml} = 6.9$
Stage 2	23	3342	2024	3802	2224	10.5	7.0	79	7.47	93	5.5	26.6	20.4	0.6	1.2	
Stage 3	24	3166	2050	3884	2248	9.0	6.7	85	7.42	92	5.6	22.5	17.4	0.7	1.4	
Stage 4	22	3272	2098	3920	2294	7.6	6.3	87	7.24	133	5.6	18.0	14.5	1.0	1.7	
Stage 5	22	3512	2360	4122	2432	11.0	5.1	87	7.20	85	5.6	15.0	13.0	1.4	2.0	
Stage 6	20	3266	2120	3962	2328	10.2	5.7	87	7.07	99	6.0	16.8	12.4	1.4	2.3	
Effluent																
From Air Train Clarifier	11	15	13	-	-	4.3	-	-	7.90	69	5.5	19.2	16.9	0.0	0.0	
From Oxygen Train Clarifier	8	9	9	-	-	4.5	-	-	7.12	74	6.0	18.0	17.1	1.0	3.6	

1 All samples were taken at the time indicated. Supernatant after 30 minutes. settling was analyzed for BOD and nutrients.

2 Units are mg./1. unless otherwise indicated

3 J.T.U. units

4 Velocity in ft/hr.

TABLE 15

INTENSIVE MIXED-LIQUOR EVALUATION

2:00 P.M. - June 24, 1969

Sample[1]	BOD[2]	TSS	VSS	TS	TVS	Turbidity[3]	Settling Velocity[4]	SVI	pH	COD	P	TKN	NH_3-N	NO_2-N	NO_3-N	Free Swim. Cilia / Stalked Cilia
Common Feed Waste	230	216	145	-	-	80.5	-	-	-	480	16.5	38.0	26.0	0.1	0.1	
Air Train																
Inlet	43	2524	1482	3614	2334	11.5	6.5	92	8.13	131	8.5	31.1	25.0	0.0	0.0	$\frac{250/ml}{90/ml} = 2.8$
Middle	20	2520	1606	3418	1884	9.8	7.1	88	7.87	100	7.7	28.0	23.1	0.0	0.0	
Exit	18	2508	1608	3438	1962	8.9	7.1	92	7.81	92	7.4	26.4	22.4	0.0	0.0	
Oxygen Train																
Stage 1	55	3128	2092	3590	2048	17.7	7.5	77	7.67	121	9.5	28.5	24.0	0.5	0.6	$\frac{310/ml}{60/ml} = 5.2$
Stage 2	60	3380	2276	3928	2318	14.2	7.1	78	7.49	237	9.0	28.5	20.6	0.5	0.7	
Stage 3	36	3138	2018	3740	2256	14.1	7.5	81	7.41	134	9.5	26.9	20.2	0.6	0.7	
Stage 4	40	2912	1954	3748	2142	15.0	7.4	87	7.23	38	9.2	21.9	20.1	1.0	1.0	
Stage 5	37	3226	2216	3998	2392	13.2	7.6	74	7.24	117	9.0	24.8	20.5	0.9	0.9	
Stage 6	39	-	-	3986	2444	16.9	7.4	-	7.20	83	8.1	26.5	20.4	1.1	0.8	
Effluent																
From Air Train Clarifier	11	13	-	-	-	2.8	-	-	-	127	6.0	23.0	22.5	0.0	0.0	
From Oxygen Train Clarifier	11	5	-	-	-	2.4	-	-	-	91	5.3	21.0	19.8	1.9	0.3	

[1] All samples were taken at the time indicated. .Supernatant after 30 minutes settling was analyzed for BOD and nutrients.

[2] Units are mg./l. unless otherwise indicated.

[3] J.T.U. units

[4] Velocity in ft/hr.

TABLE 16

INTENSIVE MIXED-LIQUOR EVALUATION

9:00 A.M. - June 25, 1969

Sample[1]	BOD[2]	TSS	VSS	Turbidity[3]	Settling[4] Velocity	SVI	pH	COD	P	TKN	NH_3-N	NO_2-N	NO_3-N	Free Swimming Cilia / Stalked Cilia
Common Feed Waste	205	173	151	-	-	-	7.92	411	6.2	30.0	24.0	0.0	0.0	
Air Train														
Inlet	40	2620	1566	7.7	6.2	97	7.82	77	4.9	28.0	19.0	0.1	0.3	$\frac{530/ml}{110/ml} = 4.8$
Middle	28	2984	1752	6.4	6.1	88	7.79	72	4.7	26.5	18.4	0.1	0.5	
Exit	21	3070	1850	5.9	6.0	98	7.77	148	6.5	21.2	18.4	0.1	0.5	
Oxygen Train														
Stage 1	60	3180	2276	18.1	7.4	75	7.36	165	7.5	28.8	16.6	0.5	0.5	$\frac{350/ml}{80/ml} = 4.4$
Stage 2	43	2588	1810	12.3	8.3	81	7.35	188	6.1	23.2	15.0	0.7	1.2	
Stage 3	39	2666	1864	10.2	7.8	87	7.27	104	5.3	18.9	13.6	1.1	1.4	
Stage 4	29	2924	1982	9.6	7.3	84	7.17	157	5.7	21.3	13.2	1.2	2.5	
Stage 5	29	3900	2820	7.6	4.5	100	7.03	119	5.8	18.3	14.0	1.4	2.7	
Stage 6	25	3650	2642	8.2	4.0	88	7.06	81	5.8	19.0	13.9	1.4	3.5	
Effluent														
From Air Clarifier	11	11	10	3.2	-	-	7.78	155	5.0	22.1	19.5	0.0	0.2	
From Oxygen Clarifier	16	10	8	4.0	-	-	6.94	63	6.0	20.9	17.9	1.4	3.6	

[1] All samples were taken at the time indicated. Supernatant after 30 minutes settling was analyzed for BOD and nutrients.

[2] Units are mg./l. unless otherwise indicated.

[3] J.T.U. units.

[4] Velocity in ft/hr.

TABLE 17

INTENSIVE MIXED-LIQUOR EVALUATION

2:00 P.M. - June 25, 1969

Sample[1]	BOD[2]	TSS	VSS	Turbidity[3]	Settling[4] Velocity	SVI	pH	COD	P	TKN	NH_3-N	NO_2-N	NO_3-N	Free Swimming Cilia / Stalked Cilia
Common Feed Waste	270	259	173	-	-	-	8.52	441	21.0	32.1	18.0	0.0	0.0	
Air Train														
Inlet	54	2487	1748	12.2	6.4	96	7.95	116	11.0	23.9	19.0	0.0	0.1	
Middle	28	2364	1730	8.6	7.5	85	7.82	81	9.1	21.9	18.4	0.0	0.0	$\frac{410/ml}{110/ml} = 3.7$
Exit	20	3088	2758	9.9	9.1	50	7.80	73	8.8	24.9	18.9	0.0	0.1	
Oxygen Train														
Stage 1	75	2428	2280	25.8	9.4	71	7.54	108	-	24.5	16.4	0.9	0.4	
Stage 2	59	2630	2184	20.4	9.5	64	7.37	143	15.7	22.6	15.2	1.1	0.6	
Stage 3	45	2356	2184	18.8	8.0	99	7.33	155	14.1	21.3	16.1	1.2	0.6	$\frac{210/ml}{60/ml} = 3.5$
Stage 4	60	3108	2888	19.5	8.0	75	7.28	110	12.4	23.5	17.1	1.3	0.9	
Stage 5	44	3116	2770	16.8	6.9	82	7.23	135	11.0	22.8	17.1	1.3	1.4	
Stage 6	47	3012	2626	16.7	7.1	85	7.13	-	9.5	23.0	17.1	1.3	1.8	
Effluent														
From Air Clarifier	14	10	9	3.6	-	-	7.90	71	6.7	22.6	19.2	0.0	0.0	
From Oxygen Clarifier	20	4	4	4.2	-	-	7.09	69	6.7	19.6	17.1	1.8	0.9	

[1] All samples were taken at the time indicated. Supernatant after 30 minutes settling was analyzed for BOD and nutrients.

[2] Units are mg./l. unless otherwise indicated.

[3] J.T.U. units.

[4] Velocity in ft/hr.

TABLE 18

INTENSIVE MIXED-LIQUOR EVALUATION

9:00 A.M. - June 26, 1969

Sample[1]	BOD[2]	TSS	VSS	Turbidity[3]	Settling Velocity[4]	SVI	pH	COD	P	TKN	NH$_3$-N	NO$_2$-N	NO$_3$-N	Free Swimming Cilia / Stalked Cilia
Common Feed Waste	245	168	124	-	-	-	8.62	251	7.3	39.5	32.1	0.0	0.1	
Air Train														
Inlet	44/50	2868	1896	6.6/2.0	6.1	98	7.92	-/49	5.8/5.4	23.8/21.8	20.5/20.1	0/0	0.1/0.1	
Middle	30/25	3278	2112	4.8/1.5	5.7	94	7.88	71/69	5.5/5.2	18.8/18.0	16.0/15.9	0/0	0.1/0.1	500/ml / 100/ml = 5.0
Exit	50/13	3402	2210	3.5/1.4	5.1	97	7.83	104/29	5.8/5.6	16.8/16.1	15.5/15.0	0/0	0.1/0.0	
Oxygen Train														
Stage 1	54/34	2666	1824	14.4/1.4	8.0	88	7.60	123/54	5.5/5.0	25.0/23.0	20.0/19.8	0.8/0.8	0.9/0.9	
Stage 2	35/27	3064	2106	11.9/1.3	7.5	80	7.58	87/42	5.2/4.8	23.2/21.6	18.0/17.5	1.1/1.1	1.5/1.3	
Stage 3	33/17	3176	2126	7.5/1.1	5.9	92	7.39	74/42	5.5/5.0	16.4/15.6	17.1/14.5	1.4/1.2	2.0/1.8	270/ml / 100/ml = 2.7
Stage 4	38/18	3804	2656	6.4/1.6	2.9	100	7.19	76/42	7.0/5.8	13.6/13.0	12.5/11.8	1.6/1.3	2.0/1.4	
Stage 5	27/15	3610	2442	8.6/1.3	5.5	86	7.10	73/41	6.5/6.2	12.8/11.3	10.8/10.9	1.6/1.3	3.3/2.9	
Stage 6	24/10	3830	2624	7.0/1.3	4.0	90	7.08	90/62	7.2/6.2	15.2/10.7	10.6/9.6	1.7/1.5	3.6/2.7	
Effluent														
From Air Clarifier	17/7	9	9	2.4/1.3	-	-	7.87	59/32	5.6/5.1	15.1/14.5	14.0/13.5	0.0/0.0	0.0/0.0	
From Oxygen Clarifier	18/6	11	11	3.4/1.1	-	-	7.10	76/67	6.7/6.4	14.6/13.8	12.5/12.2	1.5/1.4	4.1/4.0	

1 All samples were taken at the time indicated. Supernatant after 30 minutes settling was analyzed for BOD and nutrients.

2 Units are mg./l. unless otherwise indicated.

3 J.T.U. units

4 Velocity in ft/hr.

5 Where two values are indicated for BOD, Turbidity, COD and etc. Unfiltered Supernatant/Filtered Supernatant was analyzed.

Evaluation of an Oxygenation System at High Treatment Rates and High Mixed-Liquor Solids Concentration - Phase II Operation

The operation of only three stages of the six stage oxygenation unit in treatment of the entire influent raw sewage flow at the Batavia facility has been shown schematically in Figure 5. During this test period, three of the four aeration bays at the plant were taken out of service, including the entire air aerated system and one half of the oxygenation unit. This reduction in effective aerator volume from a total of 650,400 gal. (as in Phase I operation) to 162,600 gal. permitted operation at a 3 to 4 fold higher treatment rate. Both existing clarifiers were operated with the three stage oxygenation unit.

The dry weather months of July and August,1969 were selected as the most desirable time to conduct this phase of oxygenation system testing. During this period, raw sewage flows to the plant were nearly the design flow of 2.5 MGD and waste strength was high (BOD of approximately 200 mg/l or greater). Conversion of the system to this mode of operation was completed in the first two weeks of July,1969. During the third week of the month, the activated sludge level in the three stage oxygen unit was allowed to build and the biomass was acclimated to high rate treatment. Continuous data acquisition from the unit was begun in the week of July 21 and was carried out through the first week in September, 1969.

During this period of operation, all aeration equipment and process control and monitoring instrumentation functioned as planned, requiring no more than normal service maintenance. The conversion of the oxygenation system to three stage operation required only the relocation of feed oxygen and dissolved oxygen control valves. An influent sewage line already existed to transport the wastewater flow to the newly designated Stage No. 1 of the oxygen system. A recycle sludge line extention (temporary) was put in place during construction to transport return sludge to the first stage of the system. This extension passed through the first three stages of the original six stage system. A larger

(10-inch diameter) magnetic flow meter was installed in the oxygen system sludge return line replacing the existing 4-inch diameter flow meter. This was necessary to permit the recycle sludge from both clarifiers to pass through the same return line to the oxygen unit.

Sufficient excess recirculation blower capacity existed for each oxygen stage aerator to allow for the required increase in gas recirculation rates within a given stage. In general, the Stage No. 2 and No. 3 blowers having 85 cfm capacity were run at near maximum output to maintain a D.O. concentration of 8-10 mg/l. The 180 cfm blower on Stage No. 1 was operated at varying recirculation rates by automatic control to maintain the set point mixed-liquor D.O.

The system conditions predetermined for this described mode of operation were basically maintenance of an MLSS concentration of 6000-8000 mg/l (MLVSS of 4000-4500 mg/l) at a mixed-liquor D.O. concentration of 8-10 mg/l. The aeration detention time in the system would be 1.5 hours (based on raw sewage flow only) at the 2.5 MGD plant design flow. Aeration detention time including total throughput, raw flow plus recycle flow, would be 1.1 - 1.2 hours.

Examination of summary data shown in Tables 19 and 20 clearly indicate that operation of the three stage oxygenation unit at very nearly the project plan conditions was possible throughout this phase of testing. A summary of weekly averages for this period indicates a 24-hour influent raw sewage flow of approximately 2.50 MGD with a BOD concentration of approximately 220 mg/l. Mixed-liquor suspended solids concentrations in the range of 6400-7400 mg/l were maintained, the seven week average MLSS concentration being 6979 mg/l. A seven week average MLVSS concentration of 4451 mg/l was observed. The dissolved oxygen concentration in the mixed-liquor of all oxygenation stages was maintained at an average value of 9.0 mg/l for the test period. The final clarifiers were operated at an average sludge recycle/influent raw sewage flow ratio of 0.34 for the test period. The recycle sludge averaged 2.95% suspended

TABLE 19

PHASE II OPERATION - OXYGEN SYSTEM PERFORMANCE

PARAMETER	PERIOD	Weekly Average of Daily Values						
		7/21/69	7/28/69	8/4/69	8/11/69	8/18/69	9/25/69	9/1/69
Wastewater Feed Rate in MGD	Night	2.17	2.28	2.03	2.13	2.07	1.79	1.84
	Day	3.53	3.48	2.89	2.56	2.90	2.84	2.65
	24 Hour Mean	2.77	2.93	2.50	2.37	2.52	2.36	2.27
Sludge Recycle Rate as MGD	Night	0.69	0.64	0.80	0.79	1.07	0.72	0.77
	Day	0.81	0.77	0.85	0.93	1.01	0.87	0.89
	24 Hour Mean	0.73	0.71	0.83	0.87	1.04	0.80	0.83
Sludge Recycle to Raw Waste Feed Rate Ratio	Night	0.31	0.28	0.39	0.37	0.52	0.40	0.42
	Day	0.23	0.23	0.30	0.37	0.35	0.31	0.33
	24 Hour Mean	0.27	0.25	0.34	0.37	0.43	0.35	0.37
Wastewater Temp. °F	Daily Average	60	68	67	66	72	69	62
Mixed-Liquor D.O. as Average mg/l	Night	11.8	11.6	10.5	11.1	9.5	10.6	10.0
	Day	10.4	10.2	9.6	8.0	6.1	5.2	3.4
	24 Hour Mean	11.1	10.8	10.0	9.4	7.6	7.7	6.5
Clarifier Effluent D.O. in mg/l	Night	7.1	0.9	0.1	0.8	0.3	0.1	0
	Day	0.6	0.3	0.3	0.3	0.1	0.0	0
	24 Hour Mean	3.2	0.5	0.2	0.5	0.1	0.1	0
Aeration Detention Time in Hours (Wastewater & Recycle)	Night	1.4	1.3	1.4	1.4	1.2	1.6	1.5
	Day	0.9	0.9	1.1	1.1	1.0	1.1	1.1
	24 Hour Mean	1.2	1.1	1.2	1.2	1.1	1.3	1.3
Mixed-Liquor pH	Inlet Stage 1			7.8	6.9	6.7	6.9	
	Middle Stage 4			7.7	6.9	6.7	6.9	
	Outlet Stage 6			7.6	6.7	6.6	6.8	
Clarifier Overflow Rates as $Gal/Ft^2/Day$	Night	864	907	808	848	824	712	732
	Day	1405	1385	1150	1019	1154	1130	1054
	24 Hour Mean	1102	1166	995	943	1003	939	903
Nominal Aeration Detention Time in Hours (Wastewater only)	Night	1.8	1.7	1.9	1.8	1.9	2.2	2.1
	Day	1.1	1.1	1.4	1.5	1.3	1.4	1.5
	24 Hour Mean	1.4	1.3	1.6	1.6	1.5	1.6	1.7

TABLE 20

PHASE II OPERATION - OXYGEN SYSTEM PERFORMANCE

		Weekly Average of Daily Values						
PARAMETER	PERIOD	7/21/69	7/28/69	8/4/69	8/11/69	8/18/69	8/25/69	9/1/69
VSS of Feed Wastewater, mg/l	Night	104	90	79	123	83	139	65
	Day	176	125	109	171	195	157	110
	24 Hour Mean	124	107	98	134	153	150	93
Mixed Liquor TSS, mg/l	24 Hour Mean	6,464	7,134	7,388	7,348	6,759	6,535	7,227
Mixed Liquor VSS, mg/l	24 Hour Mean	3,588	4,242	4,831	4,806	4,401	4,354	4,936
Recycle Sludge TSS, mg/l	24 Hour Mean	33,942	33,714	28,511	26,787	24,782	29,594	29,599
Recycle Sludge VSS, mg/l	24 Hour Mean	18,854	20,909	18,047	17,445	16,182	18,542	17,328
Recycle Sludge VSS/TSS Ratio	24 Hour Mean	0.56	0.60	0.64	0.65	0.65	0.63	0.59
Sludge Volume Index	24 Hour Mean	39	22	34	42	41	33	41
Initial Sludge Settling Rate Ft/Hr	24 Hour Mean	7.6	8.6	6.5	4.7	5.5	7.3	5.6
Dry Solids Wasted, lbs/day	Each 24-Hours	*	2,953	1,945	2,396	2,933	2,013	*
Lbs. VSS Wasted per Lb. BOD Removed	24 Hour Mean	*	0.503	0.423	0.368	0.401	0.364	*
Food/Biomass as lb BOD/lb MLVSS	Night	0.580	0.479	0.323	0.551	0.593	0.357	0.405
	Day	1.972	0.959	0.680	0.826	1.143	0.977	0.902
	24 Hour Mean	1.344	0.705	0.519	0.704	0.887	0.693	0.675
Volumetric Organic Loading as lb BOD/Day/1000 ft^3 Mixed Liquor	Night	129.8	125.7	97.7	165.6	163.4	97.4	124.4
	Day	442.3	242.5	205.7	246.0	321.9	257.3	278.5
	24 Hour Mean	300.8	179.1	156.2	211.4	248.3	184.0	207.9
Effluent Turbidity as J.T.U.	Night		6.6	8.6	8.6	6.8	9.5	15.0
	Day		7.5	8.3	10.6	7.6	10.0	9.8
	24 Hour Mean		7.6	8.4	9.8	7.3	9.8	11.7

solids (29,561 mg/l) during the seven week period. Individual weekly average recycle sludge TSS concentrations as high as 3.4% are reported. On specific occasions, when the manually controlled sludge return pumps were carefully monitored, recycle sludge concentrations as high as 5% could readily be achieved. This recycle sludge had a light brown coloration and microscopic examination clearly indicated its viability. No odor or black to dark grey coloration indicative of sustained anaerobic conditions was evident.

As indicated in Table 21, oxygen was introduced to the three stage oxygenation unit at a 24-hour average rate of about 31 cfm throughout the seven week period. Gas was exhausted from the unit at an average rate of 4.0 cfm with an oxygen content of 55%. On an overall average basis, an oxygen utilization efficiency of 92.7% was observed. Throughout the course of the test period, this feed oxygen utilization efficiency varied by only about ± 1% from the mean. This may be taken as an indication of the highly reliable nature of the feed oxygen control system.

The overall power requirements for liquid mixing and gas recirculation in the oxygenation unit are shown in Table 22. Also shown are average power requirements per system stage on a 24-hour basis. In summary, these data indicate the distribution of power input to be an average of 11.1 HP for liquid mixing and 9.2 HP for gas recirculation within the stages. A total average power input for the seven week oxygenation system operation of 20.3 HP was observed. This power input and distribution differed from Phase I operation (26 HP for liquid mixing and 2.6 HP for gas recirculation) since more than twice the total quantity of oxygen was transferred to the mixed-liquor per day and yet three less oxygenation stages were used. The mixing energy input was thus decreased due to a smaller volume of mixed-liquor under aeration while the gas recirculation energy input was increased to meet the higher daily oxygen demand. On a unit aeration tank volume basis, the oxygenation system operated at a power input of 0.125 HP per 1000 gallons of mixed-liquor in Phase II of the study. Day to night and week to week variations in this value are listed. The power consumption

TABLE 21

PHASE II OPERATION - OXYGEN SYSTEM PERFORMANCE

PARAMETERS	PERIOD	Weekly Average of Daily Values						
		7/21/69	7/28/69	8/4/69	8/11/69	8/18/69	8/25/69	9/1/69
Total Feed Oxygen Flow Rate, CFM at NTP	Night	33.1	25.7	28.7	29.3	30.8	27.5	29.1
	Day	36.3	32.6	32.8	31.5	32.2	32.2	31.9
	24 Hour Mean	34.3	29.4	30.9	30.5	31.6	30.0	30.7
Average % Oxygen Composition in Gas Space of each Oxygen System Stage	24 Hr. Mean-Stage 1	82	82	84	82	82	83	82
	2	73	70	73	72	71	72	73
	3	52	54	58	57	54	55	56
Exhaust Gas Rate From Oxygen System 3rd Stage, CFM at NTP	Night	4.64	3.13	4.75	4.67	3.31	3.54	3.73
	Day	4.81	4.10	4.87	3.89	3.81	3.87	3.25
	24 Hour Mean	4.82	3.66	4.82	4.24	3.58	3.72	3.47
Overall % of Feed Oxygen Utilization	Night	94.8	93.2	90.6	90.6	94.0	92.7	92.6
	Day	92.0	93.2	91.2	92.9	93.5	93.4	94.2
	24 Hour Mean	92.9	93.2	91.0	91.9	93.8	93.1	93.6
Lb. Oxygen Utilized per Lb. BOD Consumed	Night	1.54	1.19	1.76	1.41	1.10	1.73	1.44
	Day	0.46	0.76	0.91	0.69	0.58	0.72	0.64
	24 Hour Mean	1.32	0.93	1.13	0.80	0.74	0.96	0.84
Oxygen Absorbed as Lb./Hr.	Night	156.0	119.0	129.1	131.8	144.2	126.8	134.7
	Day	166.1	150.9	148.6	145.7	149.5	149.3	149.6
	24 Hour Mean	158.5	136.3	139.7	139.3	147.1	139.0	142.8

TABLE 22

PHASE II OPERATION - OXYGEN SYSTEM PERFORMANCE

			Weekly Average of Daily Values					
PARAMETER	PERIOD		7/21/69	7/28/69	8/4/69	8/11/69	8/18/69	8/25/69	9/1/69
Power Required for Liquid Mixing in Oxygen System Per Stage as Hp	24 Hour Mean	Stage 1	6.0	5.9	5.9	5.6	5.8	5.9	5.5
		2	2.8	2.8	2.8	2.7	2.8	2.7	2.7
		3	3.2	3.0	2.7	2.1	2.3	2.6	2.0
Power Required for Gas Recirculation in Oxygen System per Stage as Hp	24 Hour Mean	Stage 1	4.3	3.6	2.7	4.4	4.5	4.8	6.0
		2	1.8	2.6	2.6	2.5	2.5	2.6	2.6
		3	2.7	2.4	2.5	2.4	2.4	2.5	2.4
Overall Power Required for Liquid Mixing as Hp	Night		12.0	12.3	11.4	10.7	11.0	11.3	10.5
	Day		11.9	11.2	11.3	10.3	11.1	11.1	10.0
	24 Hour Mean		11.9	11.7	11.4	10.5	11.0	11.2	10.2
Overall Power Required for Gas Recirculation as Hp	Night		6.1	7.1	6.5	8.8	8.8	8.3	11.1
	Day		10.3	9.9	8.8	9.8	9.7	11.1	11.0
	24 Hour Mean		8.7	8.6	7.7	9.4	9.3	9.8	11.1
Overall Power Utilized as Hp/1000 Gal of Mixed Liquor	Night		0.111	0.119	0.110	0.120	0.122	0.120	0.133
	Day		0.137	0.119	0.124	0.124	0.128	0.137	0.129
	24 Hour Mean		0.127	0.125	0.117	0.122	0.125	0.129	0.131

per unit mixed-liquor volume is somewhat higher than necessary in this case due to the lack of availability of the proper combination of pulleys to decrease the mixer drive speeds.

The treatment effectiveness of the high rate oxygenation system is described by the data in Tables 23, 24, and 25. On a seven week average basis, 90% of the average 220 mg/l of influent BOD was removed while the system operated at an aeration detention time of 1.2 hours (raw sewage and recycle sludge flow with a tank volume of 162,000 gallon). During the day and thus during peak organic loading, the aeration detention time averaged 1.0 hour. For the entire period of operation, a COD removal efficiency of 71% is indicated with an average influent COD concentration of 325 mg/l. The data in Table 23 also indicates that an average of 89% of influent suspended solids (174 mg/l) was removed.

These removals were achieved at the organic loading and food/biomass ratio values shown in Table 20, at averages of 212.5 lbs. BOD/day/1000 ft^3 mixed-liquor and 0.790 lbs. BOD/lb. MLVSS, respectively. On a seven week average basis, day time (10:00 a.m. - 11:00 p.m.) organic loading rates averaged 284.9 lbs. of BOD/day/1000 ft^3 of mixed-liquor and food/biomass ratios averaged 1.066 lbs. BOD/lb. MLVSS under aeration.

Listed in Table 24 are weekly average values for total phosphate analysis of feed wastewater and final clarifier effluents. Influent phosphate (as mg/l P) levels during the seven week test period averaged 12 mg/l on a 24 hour basis with comparable clarifier effluent concentrations of 10 mg/l, yielding a 12% overall removal efficiency. As in the air and oxygen comparison during the initial test period at Batavia, digester supernatant was returned to the aeration tanks. In this second phase of operation, however, the phosphorous content of the digester supernatant was not monitored in the feed waste composite sampling system. The supernatant was returned to the aeration tanks via the influent sewage line in the Phase I comparison, but limitations in piping required that this stream be returned via the sludge recycle lines in Phase II operation.

TABLE 23

PHASE II OPERATION - OXYGEN SYSTEM PERFORMANCE

		Weekly Average of Daily Values						
PARAMETER	WEEK	7/21/69	7/28/69	8/4/69	8/11/69	8/18/69	8/25/69	9/1/69
Feed Wastewater BOD,mg/1	Night	156	145	125	202	205	141	176
	Day	326	190	185	252	282	236	274
	24 Hour Mean	283	166	163	232	252	203	238
Clarifier Effluent BOD, mg/1	Night	22	19	21	20	22	24	30
	Day	31	25	22	19	29	27	18
	24 Hour Mean	29	22	22	18	26	26	23
% BOD Removed	Night	86	87	83	90	89	83	83
	Day	91	87	88	92	90	89	93
	24 Hour Mean	90	87	87	92	90	87	91
Feed Wastewater COD, mg/1	Night	243	188	205	368	257	268	424
	Day	386	308	340	442	441	335	370
	24 Hour Mean	240	265	290	405	372	312	390
Clarifier Effluent COD,mg/1	Night	114	91	102	120	72	86	103
	Day	100	108	65	123	121	61	99
	24 Hour Mean	105	100	73	122	103	70	101
% COD Removed	Night	53	52	50	67	72	68	76
	Day	74	65	81	72	73	82	73
	24 Hour Mean	68	62	75	70	72	78	74
Feed Wastewater TSS,mg/1	Night	207	96	114	164	122	144	97
	Day	323	156	143	231	331	205	131
	24 Hour Mean	209	118	133	203	253	184	119
Clarifier Effluent TSS,mg/1	Night	15	12	14	17	21	26	17
	Day	24	17	13	19	23	23	18
	24 Hour Mean	22	15	13	18	22	24	18
% TSS Removed	Night	93	88	87	90	83	82	83
	Day	93	89	91	92	93	89	86
	24 Hour Mean	89	87	90	91	91	87	85

TABLE 24

PHASE II OPERATION - OXYGEN SYSTEM PERFORMANCE

PARAMETER	PERIOD	Weekly Average of Daily Values 7/21/69	7/28/69	8/4/69	8/11/69	8/18/69	8/25/69	9/1/69
Total Phosphate in Feed Wastewater as mg/l of P	Night	7	5	8	9	9	9	8
	Day	7	12	14	16	20	16	17
	24 Hour Mean	7	10	12	13	16	13	13
Total Phosphate in Clarifier Effluent as mg/l of P	Night	7	7	9	9	9	10	10
	Day	8	10	11	10	12	13	12
	24 Hour Mean	8	9	10	9	11	12	11
Total % Phosphate Removed	Night	-5	-24	-2	-2	-4	-20	-20
	Day	-24	16	23	37	41	20	30
	24 Hour Mean	-17	8	17	26	32	11	19
Ammonia Nitrogen in Feed Wastewater as mg/l N	Night	10	9	15	17	17	15	18
	Day	13	16	22	24	31	21	25
	24 Hour Mean	12	13	19	21	26	19	23
Ammonia in Clarifier Effluent as mg/l N	Night	9	10	15	13	16	16	14
	Day	12	21	22	17	25	21	20
	24 Hour Mean	11	17	20	15	21	21	18
TKN in Feed Wastewater as mg/l N	Night	19	18	27	30	29	27	30
	Day		25	36	41	50	35	37
	24 Hour Mean		22	33	36	42	32	35
TKN in Clarifier Effluent as mg/l N	Night	11	13	20	20	23	22	18
	Day	14	23	28	24	32	26	24
	24 Hour Mean	13	20	25	22	28	25	22
NO_2 Nitrogen in Feed Wastewater as mg/l N	Night	0	0.3	0.1	0	0	0	0
	Day	0	0.0	0.0	0	0	0	0
	24 Hour Mean	0	0.1	0.0	0	0	0	0
NO_2 Nitrogen in Clarifier Effluent as mg/l N	Night	0.2	0.2	0.1	0.1	0.1	0.1	0.2
	Day	0.2	0.1	0.2	0.2	0.2	0.2	0.2
	24 Hour Mean	0.2	0.1	0.2	0.3	0.1	0.2	0.2

TABLE 25

PHASE II OPERATION - OXYGEN SYSTEM PERFORMANCE

		Weekly Average of Daily Values						
PARAMETER	PERIOD	7/21/69	7/28/69	8/4/69	8/11/69	8/18/69	8/25/69	9/1/69
NO_3 Nitrogen in Feed Wastewater as mg/l N	Night	0	0.6	0.3	0.1	0.1	0.1	0.0
	Day	0	0.1	0.1	0.1	0.2	0.1	0.1
	24 Hour Mean	0	0.3	0.2	0.1	0.2	0.1	0.1
NO_3 Nitrogen in Clarifier Effluent as mg/l N	Night	1.0	1.1	1.6	1.4	1.8	1.8	1.7
	Day	0.6	1.3	2.0	2.1	2.1	2.1	2.2
	24 Hour Mean	0.7	1.2	1.8	1.8	2.0	2.1	2.0
Total Nitrogen in Feed Wastewater as mg/l N	Night	19	19	27	30	29	28	30
	Day		25	36	41	50	35	37
	24 Hour Mean		23	33	37	42	32	35
Total Nitrogen in Clarifier Effluent as mg/l N	Night	12	14	21	21	24	23	20
	Day		25	30	27	34	29	26
	24 Hour Mean		21	27	24	30	27	24
% Total Nitrogen Removed	Night	37	23	22	29	15	16	33
	Day		2	18	35	32	18	30
	24 Hour Mean		7	18	33	28	17	31
NO_3 - N/Total N in Feed Wastewater	Night	0	0.07	0.01	0	0	0	0
	Day		0	0.00	0	0	0	0
	24 Hour Mean		0.03	0.01	0	0	0	0
NO_3 - N/Total N in Clarifier Effluent	Night	0.08	0.08	0.07	0.07	0.07	0.08	0.08
	Day		0.06	0.07	0.08	0.07	0.07	0.08
	24 Hour Mean		0.07	0.07	0.08	0.07	0.07	0.08
Lbs. Total Nitrogen Removed per Lb. of Dry Solids Wasted	24 Hour Mean	*	0.025	0.064	0.107	0.086	0.049	*

It was not then possible to arrive at a valid phosphorus balance in this phase of operation. This is noted in the seeming disparity in phosphorus removal rates between the first and second phases of operation.

In Tables 24 and 25 are listed the weekly average values for the analysis of influent wastewater and clarifier effluent for NH_3-N, TKN, NO_2-N, and NO_3-N. On an overall basis, an average of 23% of influent nitrogen was removed in the treatment process. Ammonia nitrogen concentrations of influent wastewater were on the average equivalent to Phase I operation values. TKN, however, appears to be divided nearly equally between organically bound and free ammonia nitrogen. Nitrification again occurred in the oxygenation system mixed-liquor to a limited extent. On a seven week average basis, the NO_3-N/total clarifier effluent nitrogen ratio was 0.07, indicating only 7% nitrification of effluent nitrogen. In the case of these nutrient analyses listed in Tables 24 and 25, whole rather than filtered effluent samples were used for the various determinations.

Certain of the physical characteristics of the mixed-liquor biomass are listed on a weekly average basis in Table 20. Most remarkable is the consistently low SVI of the mixed-liquor throughout the seven week period, the average value being 36. In addition to this high degree of compaction, an average initial settling velocity of 6.5 ft/hr. was observed. This settling rate is unusually high for a TSS concentration averaging 6980 mg/l in the mixed-liquor. The VSS/TSS ratio of the recycled activated sludge averaged 0.62 for the seven week test period. An average of 2448 lbs. of dry solids having this fraction of volatile solids content were wasted from the system each day. In relation to the air system dry solids wasting information from the initial phase of operation, the three stage oxygen unit assimilated nearly twice the quantity of BOD in this phase of operation yet produced less excess sludge than the air aeration system while operating at lower treatment rate. The information in Table 20 indicates that an average of 0.38 lbs. excess VSS were produced by the oxygenation system per lb. of BOD removed. The significance of this observation will be discussed.

In Table II of the Appendix to this report are listed on a weekly basis results of analyses of mixed-liquor supernatant fractions for organic and inorganic nutrients for Phase II operation. Samples from all liquid stages were taken (grab samples) and analyzed as outlined. These results serve to illustrate the effect of liquid staging on removals throughout the system. Mixed-liquor supernatant rather than filtrate was used for analysis in all cases.

Air and Oxygenation Systems Performance Comparison at High Treatment Rates and High Mixed-Liquor Solids Concentrations - Phase III Operation

In the final phase of the test program,parallel evaluation of air and oxygen aerated activated sludge treatment systems was again carried out, differing in several respects from the comparison in the initial phase operation. A diagram of the two systems has been shown in Figure 6.

During July and August, 1969, a period when the air aeration bays were out of service, temporary modifications were made on the air system in anticipation of the final phase of operation. The second in the series of two aeration bays was cleaned (10 yds^3 of silt were removed) and separated from the first air system tank by a temporary water-tight bulkhead. A recycle sludge line extension was installed so that recycle sludge might be returned to the second rather than first air aeration tank. Liquid staging baffles were installed in the second air aeration tank so that liquid staging identical to the three stage oxygenation system was accomplished. The first air aeration bay was not returned to service. These modifications made it possible in the final test period to operate parallel air and oxygen systems having the same aeration volume (162,600 gal. each) and the same liquid staging characteristics. The only major difference in the two systems was the mode of aeration. At a plant design influent flow rate of 2.5 MGD, each system would operate at an aeration detention time of approximately 3 hours without consideration of recycle sludge flow rates. By the predetermined project plan,it was intended that the oxygenation unit be operated at an MLSS concentration of 6000-8000 mg/l and a mixed-liquor D.O. concentration of 8-10 mg/l while treating one half of the influent sewage. It was intended that the air aeration system as modified should be operated at the maximum attainable MLSS level consistent with good practice (3000-4000 mg/l preferably) and at an aeration rate sufficient to maintain 2.0 mg/l of D.O. in the mixed-liquor. Fifty percent more diffusers than normal were installed in the air aeration tank and the option of tapering the areation rate along the length of the tank was possible by control of manual valves on each diffuser bank header.

Since no changes were necessary in the oxygenation unit following the second test period, it was possible to immediately begin operation of the parallel systems in the second week of September,1969. The source of seed activated sludge for the air system was the oxygenation system mixed-liquor. This sludge was permitted to acclimate for six days before continuous data taking was initiated.

With few exceptions,the process monitoring and control instrumentation functioned without difficulty throughout this test period. Continued difficulties with recycle sludge magnetic flow meters were encountered. With the high level of influent suspended solids and rags, plugging of the air system return sludge meter occurred occasionally. Grease build-up on the flow sensing elements of the recycle sludge flow meters necessitated frequent cleaning since sensitivity often varied. Two weeks of values from these meters have been excluded from the data to be presented here because of apparent inaccuracies. Additionally,the 10-inch diameter sludge return meter installed in the oxygen system during the previous phase of operation created some unusual difficulties. It was not possible to return sludge at less than 0.5 MGD rate to the oxygen system. Lack of resistance to flow in this return line decreased the head to which the air lift return pump must lift sludge in the gravity flow sink. Since each pump is capable of an 0.8 MGD rate and the performance curve is flat, attempts to decrease the pumping rate below 0.5 MGD by lowering the air rate resulted in a loss of pumping action. Decreasing the flow to the oxygenation system by partial closing of the recycle sludge line control valve caused line clogging. This annoying difficulty, unique to this set of circumstances, resulted in unusually high recycle rates and commensurately lower TSS concentration of recycle sludge. The 4-inch meter restriction in the air system sludge return line prevented this difficulty from occurring in that system.

Unusually high BOD, COD, and suspended solids concentrations of the influent sewage in October and early November made operation of the air system at high solids levels (approximately 4000 mg/l TSS) very difficult. Dissolved oxygen was limited during day time operation resulting in odor problems from the final clarifier and open sludge return sink. Maximum aeration rates for the number of diffusers present was used. Optimum performance and odor control were achieved for the air system by reducing MLSS concentrations to 3000-3500 mg/l and reducing the influent raw sewage rate by shifting the balance of the flow to the oxygenation unit which was not D.O. limited. These operational changes are reflected in the data included here.

On observation which was not quantitatively recorded is the appearance of the waste activated sludge from each system. While the oxygenation unit sludge was light brown in appearance and odor free, the air system waste sludge was visually dark and by odor appeared anaerobic. Although not quantitatively ascertained, a large portion of BOD and COD removed in the air unit appears to have been sorbed and removed from the system as waste sludge without being oxidized. This would be expected under oxygen limiting conditions.

The information in Tables 26 and 27 describe the operating conditions for the air and oxygenation systems during this nine week period of continuous data acquisition. On an overall basis, average wastewater flows of 1.29 MGD and 1.44 MGD were influent to the air and oxygenation systems,

TABLE 26

PHASE III OPERATION - AIR AND OXYGEN SYSTEMS PERFORMANCE COMPARISON

Weekly Average of Daily Values

PARAMETER	PERIOD	9/8/69		9/15/69		9/22/69		9/29/69		10/6/69		10/13/69		10/20/69		10/27/69		11/3/69	
		AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN
Wastewater Feed Rate in MGD	Night	1.04	1.01	1.31	1.03	1.27	1.19	1.04	1.00	0.99	1.16	0.97	1.21	1.12	1.38	1.11	1.53	1.20	1.81
	Day	1.32	1.59	1.51	1.50	1.49	1.54	1.35	1.35	1.22	1.53	1.42	1.58	1.39	1.54	1.53	1.73	1.72	1.89
	24 Hour Mean	1.19	1.33	1.42	1.29	1.39	1.38	1.21	1.19	1.11	1.36	1.21	1.41	1.26	1.47	1.34	1.64	1.48	1.85
Sludge Recycle Rate in MGD	Night	0.30	0.66	0.25	0.73	0.25	0.75	0.29	0.52	0.25	0.58	0.21	0.52				0.57	0.26	0.46
	Day	0.33	0.73	0.24	0.67	0.23	0.79	0.25	0.55	0.22	0.52	0.19	0.53				0.47	0.22	0.48
	24 Hour Mean	0.32	0.70	0.25	0.70	0.24	0.77	0.27	0.54	0.23	0.55	0.20	0.53				0.52	0.24	0.47
Sludge Recycle to Raw Waste Feed Rate Ratio	Night	0.31	0.71	0.20	0.73	0.20	0.65	0.29	0.55	0.25	0.51	0.24	0.50				0.37	0.22	0.25
	Day	0.27	0.46	0.16	0.45	0.16	0.58	0.19	0.42	0.22	0.35	0.14	0.34				0.27	0.13	0.25
	24 Hour Mean	0.28	0.57	0.18	0.58	0.18	0.61	0.23	0.48	0.24	0.42	0.18	0.41				0.32	0.17	0.25
Wastewater Temp. °F	Daily Avg.	66	66	73	73	72	72	72	72	71	71	70	70	67	67	68	68	68	68
Mixed Liquor D.O. as Average mg/l	Night	2.5	8.7	1.2	9.8	1.2	9.2	1.9	6.4	1.7	6.9	1.1	5.9	0.9	7.4	1.7	9.6	0.9	10.1
	Day	0	6.7	0.1	9.5	0.1	9.2	0.9	6.6	0.2	6.7	0.1	5.7	0.3	6.7	0	9.2	0	10.2
	24 Hour Mean	1.2	7.6	0.6	9.6	0.6	9.2	1.4	6.5	0.8	6.8	0.6	5.8	0.6	7.0	0.8	9.4	0.4	10.2
Clarifier Effluent D.O. in mg/l	Night	0	9.1	0	7.1	0	3.6	0	2.9	0	1.8	0	5.2	0	4.9	0	6.4	0.1	5.4
	Day	0	6.5	0	6.1	0	2.0	0	1.5	0	1.2	0	3.7	0	2.5	0	4.8	0	5.3
	24 Hour Mean	0	7.4	0	6.4	0	2.6	0	2.1	0	1.5	0	4.3	0	3.6	0	5.5	0.1	5.4
Aeration Detention Time in Hours (Wastewater & Recycle)	Night	3.0	2.5	2.5	2.2	2.6	2.0	3.0	2.7	3.2	2.3	3.4	2.3				2.0	2.7	1.8
	Day	2.4	1.7	2.2	1.8	2.3	1.7	2.5	2.1	3.0	2.0	2.4	1.9				1.8	2.0	1.7
	24 Hour Mean	2.7	2.1	2.4	2.0	2.4	1.9	2.7	2.3	3.1	2.1	2.9	2.1				1.9	2.3	1.7
Nominal Aeration Detention Time in Hours (Wastewater Only)	Night	3.7	3.8	3.0	3.8	3.1	3.3	3.7	3.9	3.9	3.4	4.0	3.2	3.1	2.8	2.9	2.4	2.6	2.2
	Day.	2.9	2.4	2.6	2.6	2.6	2.5	2.9	2.9	3.2	2.5	2.7	2.7	2.8	2.5	2.5	2.3	2.3	2.1
	24 Hour Mean	3.3	2.9	2.7	3.0	2.8	2.8	3.2	3.3	3.5	2.9	3.2	2.8	3.1	2.6	2.9	2.4	2.6	2.1
Mixed-Liquor pH	Inlet Stage 1	7.2	6.5	7.1	6.5	7.1	6.7	7.3	6.7							7.3	6.8	6.8	6.1
	Inlet Stage 2	6.9	6.4	6.9	6.4	7.2	6.5	7.3	6.6							7.3	6.8	6.7	6.0
	Inlet Stage 3	7.1	6.3	7.0	6.3	7.2	6.3	7.3	6.5							7.2	6.6	6.6	5.9

TABLE 27

PHASE III OPERATION - AIR & OXYGEN SYSTEM PERFORMANCE

PARAMETER	PERIOD	Weekly Average of Daily Values																	
		9/8/69		9/15/69		9/22/69		9/29/69		10/6/69		10/13/69		10/20/69		10/27/69		11/3/69	
		AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN
VSS of Feed Wastewater, mg/l	Night	104	104	185	185	270	270	289	289	557	557	497	497	367	367	343	343	294	294
	Day	217	217	433	433	277	277	285	285	278	278	412	412	382	382	320	320	390	390
	24 Hour Mean	175	175	339	339	274	274	300	300	370	370	468	468	376	376	339	339	346	346
Mixed Liquor TSS, mg/l	24 Hour Mean	4340	5890	4290	6810	3680	6840	3180	5890	2810	7400	3890	5700	3650	5620	3640	5560	3260	5970
Mixed Liquor VSS, mg/l	24 Hour Mean	2980	3850	3130	4300	2470	4380	2290	3750	2090	5440	2800	3760	2540	3960	2620	3960	2390	4330
Recycle Sludge TSS, mg/l	24 Hour Mean	18,530	17,960	20,210	17,700	16,680	16,260	12,650	18,760	13,540	18,870	16,670	19,490	15,770	19,060	18,580	21,500	16,810	19,580
Recycle Sludge VSS, mg/l	24 Hour Mean	12,860	11,440	13,700	12,290	11,310	10,300	9,420	12,750	9,680	13,170	12,010	13,090	11,340	13,170	13,400	14,370	12,870	13,870
Recycle Sludge VSS/TSS Ratio	24 Hour Mean	0.69	0.69	0.68	0.69	0.68	0.63	0.74	0.68	0.71	0.74	0.72	0.67	0.72	0. 9	0.72	0.67	0.77	0.71
Sludge Volume Index	24 Hour Mean	48	54	45	80	46	57	55	36	77	34	96	40	69	52	73	41	60	50
Initial Sludge Settling Rate, ft/hr	24 Hour Mean	8.5	4.4	9.4	1.7	7.9	4.2	8.3	6.7	7.2	6.3	5.8	5.9	7.3	5.1	6.4	6.0	8.4	5.4
Dry Solids Wasted, LBS	Each 24 Hours	1791	319	2943	334	4281	454	2937	371	3018	892	4842	396	3779	623	4681	1028	3177	908
Lbs. VSS Wasted per Lb. BOD Removed	24 Hour Mean	0.613	0.093	0.875	0.105	1.126	0.105	0.893	0.093	0.352	0.223	1.247	0.077	1.019	0.131	1.262	0.198	[illegible].921	0.178
Food/Biomass as 1b BOD/1b MLVSS	Night	0.394	0.274	0.391	0.200	0.751	0.406	0.855	0.383	0.966	0.464	0.595	0.453	0.797	0.517	0.650	0.689	0.549	0.456
	Day	0.713	0.777	0.722	0.528	1.115	0.526	1.190	0.724	0.900	0.341	1.110	0.905	0.782	0.573	1.092	0.842	1.235	0.763
	24 Hour Mean	0.567	0.546	0.571	0.378	0.948	0.471	1.036	0.568	0.930	0.397	0.915	0.686	0.789	0.547	0.889	0.772	0.921	0.621
Volumetric Organic Loading as 1b BOD/Day/1000 ft^3 Mixed Liquor	Night	70.0	63.2	93.1	72.9	114.3	106.0	101.8	98.8	123.9	149.2	87.1	130.1	122.4	148.6	107.5	155.7	89.7	133.9
	Day	138.7	172.3	146.3	145.6	150.1	152.3	161.3	158.9	108.5	136.9	178.6	202.5	145.6	160.8	162.7	182.8	189.4	208.4
	24 Hour Mean	107.2	122.3	121.0	110.7	133.7	131.1	130.1	127.5	115.5	142.6	136.9	169.4	135.0	155.2	137.4	170.4	143.8	174.4
Effluent Turbidity as J.T.U.	Night	9.5	8.9	5.4	6.7	3.1	3.3	6.4	7.6	4.8	2.6	4.1	1.3	3.3	2.1	4.9	3.8	6.7	4.1
	Day	8.3	9.3	5.5	6.3	3.8	2.5	6.8	5.4	6.1	4.6	4.7	2.0	4.7	2.6	5.4	3.0	8.4	3.1
	24 Hour Mean	8.8	9.2	5.4	6.4	3.5	2.8	6.6	6.2	5.5	3.8	4.5	1.7	4.2	2.4	5.2	3.3	7.7	3.8

respectively. Aeration detention times (raw sewage and recycle sludge flow) for each system, air and oxygenation, were approximately 2.4 and 2.0 hours, respectively, on an average basis. Mixed-liquor suspended solids concentrations were maintained at an average concentration of 6185 mg/l in the oxygenation system and 3637 mg/l in the air aeration system. The corresponding average MLVSS concentrations were 4305 mg/l and 2576 mg/l, respectively. During the nine week period, average dissolved oxygen concentrations of 0.8 mg/l and 8.0 mg/l were maintained in the respective air and oxygenation systems mixed-liquor. Clarifier effluent D.O. for the oxygenation system averaged 4.3 mg/l during the nine week period while that of the air aerated system averaged 0.1 mg/l. Recycle sludge concentrations as TSS averaged 16,600 mg/l for the air system and 18,790 mg/l for the oxygenation unit. The recycle sludge flow rate/raw sewage influent flow rate averaged 0.21 for the air aerated treatment unit and 0.45 for the oxygen aerated unit. The reason for the disparity between the two ratios has been indicated previously. Clarifier overflow rates averaged 1027 and 1142 gal/ft^2/day for the air and oxygenation treatment systems, respectively.

Listed in Table 28 are weekly average values for the air aeration rates used on both a 24-hour and day-night basis. Over the nine week operation period, an average air aeration rate of 4.32 ft^3/fal of sewage treated was used. This required a net input of 138.5 air blower HP. As indicated in Table 28, pure oxygen was fed to the multistage oxygenation system at an average rate of 19.0 cfm during the nine week test period. Gas was exhausted from the final stage of the oxygenation system at an average rate of 3.22 cfm containing 51% oxygen. The overall average oxygen utilization efficiency for the multistage system was 91.4%. Also shown in Table 28 are weekly average values for lbs. oxygen utilized/lb. BOD removed, an average of 0.72 lbs. oxygen/lb. BOD for the entire test period.

Listed in Table 29 is the weekly average power input to each stage of the oxygenation system in terms of HP required for liquid mixing and aeration gas recirculation to transfer the lbs. of oxygen/hour noted in

TABLE 28

PHASE III OPERATION - AIR & OXYGEN SYSTEM PERFORMANCE

PARAMETER	PERIOD	Weekly Average of Daily Values																	
		9/8/69		9/15/69		9/22/69		9/29/69		10/6/69		10/13/69		10/20/69		10/27/69		11/3/69	
		AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN
Ft^3 Air Utilized per Gallon Sewage Treated, NTP	Night	5.69		4.22		4.38		5.48		5.57		5.57		5.21		5.07		4.43	
	Day	4.12		3.69		3.66		4.05		4.59		3.38		4.87		3.36		2.99	
	24 Hour Mean	4.75		3.91		3.96		4.62		4.99		4.18		5.01		4.01		3.52	
Air Blower Power Utilized as Hp	Night	153.2		142.4		140.0		143.2		138.8		133.3		144.4		136.6		129.1	
	Day	139.9		143.2		137.3		138.1		140.7		118.6		167.1		125.1		124.9	
	24 Hour Mean	145.9		142.8		138.6		140.4		139.9		125.3		156.7		130.4		126.8	
Total Feed Oxygen Flow Rate, CFM at NTP	Night		17.9		16.3		17.2		18.5		13.0		18.1		18.9		17.4		17.2
	Day		21.9		18.4		19.7		21.3		19.4		22.7		21.3		20.2		19.2
	24 Hour Mean		20.1		17.4		18.6		20.0		16.5		20.6		20.2		18.9		18.3
Average % Oxygen Composition in Gas Space of each Oxygen System Stage	24 Hour Mean Stage 1		78		78		80		80		75		80		80		82		81
	2		67		67		69		71		66		70		68		67		67
	3		51		50		54		53		46		47		50		53		53
Exhaust Gas Rate from Oxygen System Final Stage	Night		3.14		2.08		2.70		2.99		1.68		4.48		5.18		4.19		3.26
	Day		2.26		2.07		2.25		3.28		2.10		4.90		4.00		4.31		3.26
	24 Hour Mean		2.66		2.08		2.46		3.15		1.91		4.70		4.54		4.25		3.26
Overall % of Feed Oxygen Utilized	Night		90.5		93.7		91.7		91.8		94.0		88.1		86.1		87.7		90.1
	Day		94.6		94.1		93.6		91.5		94.7		90.2		89.0		88.4		91.1
	24 Hour Mean		93.0		94.0		92.8		91.6		94.4		89.6		87.9		88.2		90.7
LB Oxygen Utilized Per lb. BOD Consumed	Night		1.58		1.28		0.87		1.03		0.47		0.76		0.63		0.59		0.69
	Day		0.73		0.71		0.69		0.72		0.78		0.59		0.68		0.56		0.48
	24 Hour Mean		0.93		0.90		0.76		0.84		0.63		0.64		0.66		0.57		0.55
Oxygen Absorbed as lb/hr	Night		80.8		76.0		78.5		84.2		60.9		79.9		81.2		76.0		77.1
	Day		102.8		86.2		91.8		97.1		91.9		102.0		95.6		88.5		86.9
	24 Hour Mean		92.7		81.5		85.7		91.2		77.7		91.9		89.0		82.8		82.4

TABLE 29

PHASE III OPERATION - AIR & OXYGEN SYSTEM PERFORMANCE

PARAMETER	PERIOD	Weekly Average of Daily Values																	
		9/8/69		9/15/69		9/22/69		9/29/69		10/6/69		10/13/69		10/20/69		10/27/69		11/3/69	
		AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN
Power Required for Liquid Mixing in Oxygen System per Stage as Hp	24 Hour Mean Stage 1		5.7		5.7		5.4		5.7		5.6		4.8		5.4		4.6		4.9
	2		2.6		2.7		2.7		2.8		2.9		2.9		2.9		2.9		2.9
	3		3.1		3.1		3.4		3.3		3.7		4.0		4.0		4.1		3.8
Power Required for Gas Recirculation in Oxygen System Per Stage as Hp	24 Hour Mean Stage 1		5.2		4.7		4.6		4.9		5.9		4.8		6.4		6.8		7.4
	2		2.5		2.4		1.7		2.2		2.1		3.0		2.8		2.5		2.5
	3		2.5		2.6		2.4		2.6		2.6		2.6		2.6		2.7		2.9
Overall Power Required for Liquid Mixing as Hp	Night		11.4		11.7		11.4		11.7		12.1		11.2		12.0		11.3		11.4
	Day		11.3		11.4		11.6		11.7		12.3		12.0		12.6		12.0		11.7
	24 Hour Mean		11.4		11.5		11.5		11.7		12.2		11.6		12.3		11.7		11.6
Overall Power Required for Gas Recirculation as Hp	Night		10.2		9.5		9.3		8.3		11.0		10.7		11.1		11.7		12.8
	Day		10.1		9.8		8.3		10.0		10.1		10.1		12.4		12.1		12.8
	24 Hour Mean		10.1		9.7		8.8		9.6		10.5		10.4		11.8		11.9		12.8
Overall Power Utilized as Hp/1000 Gal. of Mixed Liquor	Night		0.133		0.130		0.128		0.123		0.142		0.135		0.142		0.141		0.148
	Day		0.132		0.130		0.122		0.133		0.137		0.136		0.154		0.148		0.151
	24 Hour Mean		0.132		0.130		0.125		0.131		0.140		0.135		0.148		0.143		0.150
Clarifier Overflow Rates as Gal/Ft^2/Day	Night	828	804	1043	820	1011	947	828	796	787	923	771	963	891	1098	883	1218	955	1440
	Day	1050	1265	1202	1194	1186	1226	1074	1074	971	1218	1130	1257	1106	1226	1218	1377	1369	1504
	24 Hour Mean	947	1058	1130	1027	1106	1058	963	947	883	1082	963	1122	1003	1170	1067	1305	1178	1472

Table 28. On an overall nine week average basis, power input of 10.6 HP was required for gas recirculation and 11.7 HP for liquid mixing, a total system power input average of 22.3 HP. On a unit volume basis, an average of 0.137 HP/1000 gal. of mixed-liquor was applied to the system.

In Table 30 are listed weekly average values for treatment effectiveness with respect to BOD, COD and influent TSS by both the air and oxygenation systems. On an overall average basis, an influent BOD concentration of 262 mg/l was observed for both the air and oxygenation systems. Overall average BOD removals of 88% and 94% were attained in the respective treatment systems, corresponding to average effluent BOD concentrations of 30 mg/l and 14 mg/l. Unusually high concentrations of BOD and COD were found in the influent sewage during September and October associated with correspondingly high concentration of TSS. This increased organic load appeared to be constituted by winery or canning wastes, perhaps both. A portion of the influent solids were floating, fibrous material. Since no primary treatment system existed at the plant, these materials were fed directly to the aeration systems.

COD concentrations of the influent waste average 578 mg/l during the nine week period. Removal efficiencies of 79% and 84% were determined for the air and oxygenation systems, respectively. Clarifier effluent COD concentrations averaging 116 mg/l and 89 mg/l were determined for the air and oxygenation systems. The average TSS concentration of influent sewage was 430 mg/l of which 94% was removed in the air aeration system and 97% in the oxygenation treatment system.

The values in Table 31 indicate the weekly average results of total phosphorus analyses of the influent wastewater and final clarifier effluents from each of the treatment units. Of an average influent phosphorus concentration of 19 mg/l for the nine week operation period, 32% was removed by the air aerated system and 27% by the oxygenation treatment system.

TABLE 30

PHASE III OPERATION - AIR & OXYGEN SYSTEM PERFORMANCE

Weekly Average of Daily Values

PARAMETER	PERIOD	9/8/69		9/15/69		9/22/69		9/29/69		10/6/69		10/13/69		10/20/69		10/27/69		11/3/69	
		AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN
Feed Wastewater BOD,mg/l	Night	169	169	182	182	233	233	256	256	329	329	259	259	282	282	259	259	193	193
	Day	278	278	252	252	260	260	308	308	232	232	331	331	273	273	275	275	289	289
	24 Hour Mean	237	237	221	221	249	249	283	283	270	270	304	304	277	277	269	269	249	249
Clarifier Effluent BOD,mg/l	Night	33	23	24	17	25	14	36	22	30	8	24	9	24	11	27	14	30	15
	Day	35	21	31	19	27	14	46	17	25	9	28	12	23	10	33	15	37	12
	24 Hour Mean	34	22	28	18	26	11	42	19	27	9	27	11	23	10	30	15	34	14
% BOD Removed	Night	80	86	87	91	89	94	86	92	91	98	91	96	91	96	90	94	84	92
	Day	87	92	88	92	90	95	85	94	89	96	91	96	92	96	88	95	87	96
	24 Hour Mean	86	91	87	92	90	95	85	93	90	97	91	96	92	96	89	95	86	94
Feed Wastewater COD,mg/l	Night	301	301	406	406	524	524	494	494	631	631	631	631	810	810	708	708	583	583
	Day	507	507	624	624	576	576	550	550	569	569	705	705	533	533	639	639	595	595
	24 Hour Mean	430	430	521	521	555	555	528	528	583	583	677	677	649	649	667	667	590	590
Clarifier Effluent COD,mg/l	Night	120	115	129	116	111	80	126	78	105	76	92	64	104	62	138	63	132	74
	Day	114	102	113	128	93	107	161	64	96	117	100	108	101	83	118	100	138	61
	24 Hour Mean	116	106	120	114	101	96	147	69	99	101	97	91	102	74	126	84	135	67
% COD Removed	Night	60	62	68	72	79	85	74	84	83	88	85	90	87	92	80	91	77	87
	Day	77	80	82	79	84	81	71	88	83	80	86	85	81	84	82	84	77	90
	24 Hour Mean	73	75	77	78	82	83	72	87	83	83	86	87	84	89	81	87	77	89
Feed Wastewater TSS,mg/l	Night	130	130	229	229	366	366	439	439	617	617	643	643	425	425	438	438	370	370
	Day	291	291	585	585	387	387	384	384	345	345	546	546	458	458	418	418	502	502
	24 Hour Mean	231	231	453	453	378	378	421	421	453	453	612	612	444	444	429	429	440	440
Clarifier Effluent TSS,mg/l	Night	29	14	14	12	15	14	23	16	34	11	14	7	14	13	19	13	21	11
	Day	28	13	17	10	16	10	37	17	25	14	22	9	18	12	25	11	29	9
	24 Hour Mean	28	13	15	10	16	12	31	17	30	13	19	8	17	12	23	12	26	10
% TSS Removed	Night	78	89	94	95	96	96	95	96	94	98	98	99	97	97	96	97	94	97
	Day	90	96	97	98	96	97	90	96	93	96	96	98	96	97	94	97	94	98
	24 Hour Mean	88	94	97	98	96	97	93	96	93	97	97	99	96	97	95	97	94	98

TABLE 31

PHASE III OPERATION - AIR & OXYGEN SYSTEM PERFORMANCE

Weekly Average of Daily Values

PARAMETER	PERIOD	9/8/69		9/15/69		9/22/69		9/29/69		10/6/69		10/13/69		10/20/69		10/27/69		11/3/69	
		AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN
Phosphate in Feed Wastewater as mg/l of P	Night	10	10	11	11	16	16	14	14	21	21	28	28	22	22	21	21	16	16
	Day	18	18	23	23	18	18	20	20	20	20	26	26	21	21	21	21	18	18
	24 Hour Mean	15	15	18	18	17	17	17	17	21	21	26	26	22	22	21	21	18	18
Phosphate in Clarifier Effluent as mg/l of P	Night	11	12	11	13	12	15	11	14	15	15	14	15	14	14	14	14	12	13
	Day	13	14	12	13	13	16	13	14	14	15	14	15	13	13	15	14	12	11
	24 Hour Mean	12	13	12	13	12	15	12	14	14	15	14	15	13	13	15	14	12	12
% Phosphate Removed	Night	-16	-24	-4	-24	26	7	24	1	29	28	50	48	39	37	31	34	24	23
	Day	30	25	46	40	31	14	35	32	30	23	45	43	38	38	31	34	34	41
	24 Hour Mean	19	13	34	24	29	11	31	21	33	27	47	45	38	38	31	34	30	33
Ammonia Nitrogen in Feed Wastewater as mg/l N	Night	26	26	20	20	28	28	32	32	38	38	36	36	35	35	36	36	27	27
	Day	23	23	27	27	27	27	30	30	28	28	33	33	31	31	41	41	28	28
	24 Hour Mean	24	24	24	24	28	28	31	31	31	31	34	34	33	33	39	39	28	28
Ammonia in Clarifier Effluent as mg/l N	Night	24	25	18	25	22	25	23	24	27	26	27	24	27	26	26	28	20	21
	Day	20	22	23	25	24	25	23	25	21	25	27	24	24	25	28	26	22	19
	24 Hour Mean	22	23	21	25	24	25	23	24	24	25	27	24	26	25	27	27	21	20
TKN in Feed Wastewater as mg/l N	Night	38	38	34	34	52	52	52	52	63	63	70	70	73	73	65	65	47	47
	Day	38	38	54	54	48	48	50	50	58	58	64	64	52	52	54	54	47	47
	24 Hour Mean	38	38	47	47	49	49	51	51	58	58	66	66	61	61	58	58	47	47
TKN of Clarifier Effluent as mg/l N	Night	31	31	25	30	27	31	33	29	35	32	33	31	35	34	31	32	27	25
	Day	27	26	29	29	28	31	35	31	27	34	34	32	34	30	34	30	30	20
	24 Hour Mean	28	28	27	29	28	31	34	30	30	33	34	32	34	31	33	31	28	22
NO_2 Nitrogen in Feed Wastewater as mg/l N	Night	0.2	0.2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
	Day	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
	24 Hour Mean	0.1	0.1	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO_2 Nitrogen in Clarifier Effluent as mg/l N	Night	0.1	0.1	0.1	0.2	0.1	0.7	0.2	1.3	0.2	0.5	0.1	0.6	0.1	0.6	0.6	1.2	0.2	1.4
	Day	0.1	0.1	0.2	0.2	0.1	0.5	0.3	1.0	0.6	0.2	0.1	0.5	0.2	0.7	0.3	1.2	0.3	1.6
	24 Hour Mean	0.1	0.1	0.2	0.2	0.1	0.6	0.3	1.1	0.5	0.3	0.1	0.5	0.2	0.7	0.4	1.2	0.3	1.5

In Tables 31 and 32 are weekly average results of analyses for influent and effluent NH_3-N, TKN, NO_2-N and NO_3-N. Removals of ammonia and TKN for each system were essentially equivalent. An overall average removal efficiency of approximately 37% was observed in both instances. The analysis for effluent nitrate nitrogen in both systems indicates an overall average concentration of 2.0 mg/l. At this limited level of nitrification,the average NO_3-N/total N ratio in the clarifier effluents of 0.07 indicates a 7% nitrification efficiency in each system.

The settling characteristics of the biomass under aeration in both air and oxygenation systems are listed in Table 27. On an overall average basis, the mixed-liquor SVI values for the air and oxygenation systems were 63 and 49,respectively. Comparative initial settling velocities are also indicated. As shown, the air aeration system was operated at an organic loading averaging 129 lbs. BOD/Day/1000 cf of mixed-liquor with a food/biomass ratio of 0.841 lbs. of BOD/lb. MLVSS. The oxygenation system organic loading averaged 145 lbs. BOD/Day/1000 cf of mixed-liquor with an average food/biomass ratio of 0.554 lbs. BOD/lb. MLVSS. Also listed on a weekly average basis in Table 27 are the comparative amounts of activated sludge wasted from each system. On an overall average basis for the nine week test period, 3490 lbs. of dry solids (71% volatile solids) were wasted each day from the air aeration system. At the same time only 600 lbs. of solids (69% volatile solids) were wasted from the oxygenation unit. This decreased level of excess solids production by the oxygenation system is striking when it is considered that over 50% of the BOD, COD, and suspended solids influent to the plant were treated in this system. On an average comparative basis, 0.989 lbs. of excess VSS were produced in the air system per lb. of BOD removed while in the oxygenation system 0.134 lbs. of excess VSS were produced per lb. of BOD removed. The proposed explanation for this dramatic reduction in excess solids in the oxygenation unit will be discussed in detail.

In Table III of the Appendix are listed certain weekly analyses of mixed-liquor supernatant fractions grab-sampled at the same time from each

TABLE 32

PHASE III OPERATION - AIR & OXYGEN SYSTEM PERFORMANCE

Weekly Average of Daily Values

PARAMETER	PERIOD	9/8/69		9/15/69		9/22/69		9/29/69		10/6/69		10/13/69		10/20/69		10/27/69		11/3/69	
		AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN	AIR	OXYGEN
NO_3 Nitrogen in	Night	0.1	0.1	0.0	0.0	0.1	0.1	0.2	0.2	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1
Feed Wastewater	Day	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1
as mg/l N	24 Hour Mean	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1
NO_3 Nitrogen in	Night	1.7	1.2	2.2	2.0	3.6	2.5	2.4	3.3	3.2	2.8	2.0	1.6	1.4	1.5	1.6	2.0	1.7	2.0
Clarifier Effluent	Day	1.4	1.2	2.4	2.0	3.3	2.2	2.8	2.9	3.9	2.5	1.8	1.6	1.4	1.5	2.1	2.2	1.4	2.3
as mg/l N	24 Hour Mean	1.5	1.2	2.3	2.0	3.4	2.3	2.7	3.0	3.6	2.7	1.9	1.6	1.4	1.5	1.9	2.1	1.5	2.2
Total Nitrogen in	Night	38	38	34	34	52	52	52	52	63	63	70	70	73	73	65	65	47	47
Feed Wastewater	Day	38	38	55	55	48	48	50	50	58	58	64	64	52	52	54	54	47	47
as mg/l N	24 Hour Mean	38	38	47	47	50	50	51	51	58	58	66	66	61	61	58	58	47	47
Total Nitrogen in	Night	32	32	27	32	31	34	35	34	38	36	35	33	36	36	33	35	29	25
Clarifier Effluent	Day	28	27	32	32	32	33	38	35	31	37	36	34	35	32	36	33	31	24
as mg/l N	24 Hour Mean	30	29	30	32	31	34	37	35	34	36	36	34	36	34	35	34	30	25
% Total Nitrogen	Night	15	15	20	7	41	34	32	34	39	44	50	53	50	51	49	46	40	47
Removed	Day	26	28	42	42	34	31	25	31	46	37	44	46	32	39	33	38	34	49
	24 Hour Mean	22	24	36	32	37	32	28	32	41	38	46	49	42	45	40	42	36	48
NO_3 - N/Total N	Night	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
in Feed Wastewater	Day	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
	24 Hour Mean	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO_3 - N/Total N	Night	0.05	0.04	0.08	0.06	0.12	0.07	0.07	0.10	0.09	0.08	0.06	0.05	0.04	0.04	0.05	0.05	0.06	0.15
in Clarifier Effluent	Day	0.05	0.04	0.08	0.06	0.11	0.06	0.08	0.08	0.14	0.07	0.05	0.05	0.04	0.05	0.06	0.07	0.05	0.10
	24 Hour Mean	0.05	0.04	0.08	0.06	0.11	0.07	0.07	0.09	0.12	0.07	0.05	0.05	0.04	0.05	0.05	0.06	0.05	0.12
Lbs. Total Nitrogen Removed per lb. of Dry Solids Wasted	24 Hour Mean	0.044	0.313	0.068	0.483	0.051	0.405	0.048	0.428	0.074	0.279	0.063	0.950	0.069	0.531	0.055	0.319	0.066	0.374

liquid stage of the two areation systems during Phase III operation. These values are provided to illustrate the effects of the parallel liquid staging on removal of the organic and inorganic nutirents in the system.

DISCUSSION

Oxygenation System Equipment Performance and Reliability

The multistage oxygenation system installed at the Batavia Municipal Treatment Plant has been described earlier in this report. This unit was the first of its kind to be installed and operated on a full scale basis. Any discussion of system reliability must, therefore, take into account the temporary nature of the installation and the flexibility required by the variable test program for its operation. With this point considered, it can be stated that the multistage oxygenation unit functioned in a highly reliable manner during the periods of operation described here. Maintenance and service for the system components were not excessive and no major modifications were required after the initial installation. As has been described elsewhere, minor modifications were necessary to enable the oxygenation system to function differently under the widely varying modes of operation in each phase of the planned test program.

In the absence of technicians assigned to the project, the oxygenation unit was attended and operated by regular treatment plant personnel, except during the period from 11:00 p.m. to 7:00 a.m. each day when there was no operator on duty. This illustrates that the multistage unit did not require unusual attention for reliable operation. Control of feed oxygen gas to the system was fully automatic as was the control of mixed-liquor dissolved oxygen concentrations. Attention to the oxygen aeration equipment by plant operators was necessary only in the event of electrical power failure. In this event, no special knowledge was required to restart the turbine drives and gas recirculation blowers. Clarifier operation and sludge wasting were manually controlled in the normal manner for both the air and oxygenation systems. Plant operator duties were, therefore, not unlike those for any conventionally operated air aeration system. No special changes were made in the basic hydraulic function of the existing plant other than the minor modifications necessitated for operating two dissimilar processes in parallel as required by the project plan.

Mixing and gas recirculation energy utilized to operate the system confirmed energy requirement predictions based on earlier pilot plant studies. Thus, the expected high oxygen mass transfer efficiencies were achieved. The mixing energy applied in each system stage, although much lower than that applied by necessity to a conventional diffused air system, was wholly adequate for maintaining a homogenous distribution of mixed-liquor solids and dissolved oxygen. Underwater inspection revealed no accumulation of rags and other debris on the impeller or rotating sparger. The homogenous production and distribution of exceedingly fine bubbles from the gas-liquid contacting units were obtained as expected from earlier pilot studies.

Underwater inspection of baffles and gas-liquid contacting equipment during the program and after shutdown of the unit at the conclusion of the contract period revealed no unusual corrosion or other deterioration of carbon steel components. No buildup of sediment on the floor of the oxygenation system stages was observed.

The reliability of the feed oxygen control system in response to shock loads and diurnal load variations was excellent. This is verified in the Results by the consistent maintenance of high D.O. concentration at all times in the mixed-liquor while treatment efficiencies remained consistently high. The gas tight covers on the oxygenation system stages presented no difficulty in containing the gas of high oxygen composition in the system. This would be expected since positive pressure under these covers never exceeded three inches of water column. As evidenced by visual examination, bubble entrainment and subsequent gas loss in the effluent mixed-liquor from the final stage of the oxygenation system was completely avoided by the use of an overflow weir to the mixed-liquor effluent line.

This weir also prevented marked changes in liquid level in the multistage unit. As a result, changes in gas space pressure under the tank covers resulting from hydraulic surges were negligible.

Realizing that the multistage oxygenation treatment system functions by utilizing an aeration gas having a high percent oxygen composition relative to air, it is appropriate to comment on the safety considerations involved in the operation of the unit. The gas separation industry has had extensive experience in the design of equipment and evaluation of materials which may be used with confidence in atmospheres having high oxygen concentrations. The result of this background technology is evident in the unattended operation of many on-site oxygen production facilities throughout the nation. The same background technology has been employed in the selection of equipment and materials used in the multistage oxygenation system tested at Batavia. This factor along with the methods (sealed covers) used to contain the oxygen aerating gas within the system eliminated a requirement for any unusual precautions. The equipment and materials within the system do not support combustion under the operating conditions described. Although not observed during the test work described here, the possible introduction of low molecular weight hydrocarbons to the oxygenation system should be considered. In general, only trace quantities of volatile hydrocarbons may be dissolved in feed wastewater to a domestic treatment plant because of the extremely low solubility of these materials in water. As a result of this low solubility, dissolved hydrocarbons may be stripped from wastewater in normal primary treatment processing in either the aerated grit removal chamber or in the primary clarifier as the effluent flows over the weirs. In the more unusual instance (heavy spills), where floating liquid hydrocarbons might be contained in the feed wastewater, complete removal in the primary treatment may not be possible. For this reason, the oxygenation system may be equipped with a hydrocarbon monitor so that oxygen flow to the unit is automatically terminated before a hazardous condition occurs.

Overall Treatment Effectiveness

A detailed listing of operating and performance data obtained in the experimental evaluation and comparison of the oxygenation treatment system with conventional air aeration treatment has been given in the Results. This week by week listing of daily average performance data is the synthesis of over 100,000 individual observations and analyses. As such, many conclusions are apparent by direct examination of these data and require no further interpretation. For this reason, the discussion of the relative performance characteristics of the oxygenation system will serve to put certain overall considerations in their proper perspective. As an aid in reference to the overall conclusions drawn from this work, a summary contrasting average operating conditions and systems performance is shown here in Tables 33, 34, 35, and 36.

The operating flexibility designed into the Batavia treatment plant permitted evaluation and comparison of the oxygenation treatment system under three distinctly different sets of conditions. The detailed description of the means used to accomplish each set of conditions has been described. In Phase I operation, conditions for evaluation of the oxygenation system were quite conservative in terms of ultimate system performance capability. Treatment detention times were those of a more conventionally operated air aerated system while mixed-liquor solids concentrations (MLSS and MLVSS), food to biomass ratios (lbs. BOD/day/lb. MLVSS) and organic loadings (lbs. BOD/day/1000 ft^3 mixed-liquor) were not unusual. By comparison, the oxygenation system differed from the parallel air system in the high aerating gas oxygen partial pressure, low mixing energy input, consistently high mixed-liquor dissolved oxygen concentration and staging of the mixed-liquor and aerating gas. The only difference in operation of the air system from conventional designed conditions was the rate of aeration, about 3.0 ft^3 of air/gal. of waste treated compared with a design for about 1.8 ft^3/gal. Major areas of difference in the comparison were then aerating gas oxygen composition, mixed-liquor D.O. sustained, liquid staging and level of energy input to transfer oxygen.

TABLE 33

COMPARATIVE SUMMARY OF AIR AND OXYGENATION TREATMENT SYSTEM PERFORMANCE

	Phase I Operation		Phase II Operation	Phase III Operation	
	Air System	Oxygenation System	Oxygenation System	Air System	Oxygenation System
Wastewater Feed Rate (MGD)	1.97	1.91	2.53	1.29	1.44
Sludge Recycle Rate (MGD)	0.25	0.44	0.83	0.25	0.60
Recycle Sludge/Wastewater Feed Rate Ratio	0.13	0.24	0.34	0.21	0.45
Wastewater Temperature (°F)	59	59	66	70	70
Mixed-liquor average D.O. Conc.(mg/1)	1.5	8.7	9.0	0.8	8.0
Clarifier Effluent D.O. Conc. (mg/1)	0.4	4.0	0.7	0	4.3
*Aeration Detention Time (hrs.)	3.5	3.4	1.2	2.6	2.0
**Nominal Aeration Detention Time (hrs.)	4.0	4.1	1.5	3.0	2.8
MLSS Concentration (mg/1)	2440	3060	6980	3640	6190
MLVSS Concentration (mg/1)	1740	2210	4450	2580	4310
Recycle Sludge TSS (mg/1)	14,960	18,620	29,560	16,600	18,790
Recycle Sludge VSS (mg/1)	9,710	11,760	18,180	11,840	12,720
Recycle Sludge VSS/TSS Ratio	0.65	0.63	0.62	0.71	0.69
Mixed-liquor Sludge Volume Index	76	64	36	63	49
Mixed-liquor Initial Settling Velocity, (ft/hr)	7.7	7.2	6.5	7.7	5.1
Final Clarifier Overflow Rate as gal/ft^2/day	1570	1520	1010	1030	1140

* Raw flow plus recycle sludge flow.

** Raw wastewater flow only.

TABLE 34

COMPARATIVE SUMMARY OF AIR AND OXYGENATION TREATMENT SYSTEMS PERFORMANCE

	Phase I Operation		Phase II Operation	Phase III Operation	
	Air System	Oxygenation System	Oxygenation System	Air System	Oxygenation System
Feed Wastewater BOD (mg/1)	159	159	220	262	262
Clarifier Effluent BOD (mg/1)	16	11	23	30	14
% BOD Removal	90	92	90	88	94
Feed Wastewater COD (mg/1)	352	352	325	578	578
Clarifier Effluent COD(mg/1)	84	73	97	116	89
% COD Removal	76	80	71	79	84
Feed Wastewater TSS (mg/1)	221	221	174	430	430
Clarifier Effluent TSS (mg/1)	16	9	19	23	12
% TSS Removal	93	96	89	94	97
Feed Wastewater VSS (mg/1)	152	152	123	332	332
Effluent Turbidity (J.T.U.)	5.7	5.6	9.1	5.7	4.4
Food/Biomass Ratio as lbs.BOD/Day/lb. MLVSS	0.566	0.414	0.790	0.841	0.554
Volumetric Organic Loading as lbs.BOD/day/1000 ft^3 mixed-liquor	60.0	57.9	212.5	128.9	144.8
Lbs. Dry Solids Wasted/Day	3148	1804	2448	3494	592
Lbs. VSS Wasted per Lb. BOD Removed	0.871	0.482	0.412	0.989	0.134

TABLE 35

COMPARATIVE SUMMARY OF AIR AND OXYGENATION TREATMENT SYSTEMS PERFORMANCE

	Phase I Operation		Phase II Operation	Phase III Operation	
	Air System	Oxygenation System	Oxygenation System	Air System	Oxygenation System
Feed Wastewater Total Phosphate(mg/1)	11	11	12	19	19
Clarifier Effluent Total Phosphate (mg/1)	6	7	10	13	14
% Total Phosphate Removed	44	39	12	32	27
Feed Wastewater NH_3-N (mg/1)	20	20	19	30	30
Clarifier Effluent NH_3-N (mg/1)	17	15	18	24	24
Feed Wastewater TKN (mg/1)	34	34	33	53	53
Clarifier Effluent TKN (mg/1)	22	19	22	31	30
Feed Wastewater NO_2-N (mg/1)	0.2	0.2	0	0	0
Clariffer Effluent NO_2-N(mg/1)	0.4	0.9	0.2	0.2	0.7
Feed Wastewater NO_3-N (mg/1)	0.3	0.3	0.1	0.1	0.1
Clarifier Effluent NO_3-N (mg/1)	0.6	1.6	1.7	2.2	2.5
Feed Wastewater Total Nitrogen (mg/1)	35	35	34	53	53
Clarifier Effluent Total Nitrogen(mg/1)	23	21	26	33	33
% Total Nitrogen Removed	35	37	23	36	36
NO_3-Nitrogen/Total Nitrogen in Clarifier Effluent	0.03	0.08	0.07	0.07	0.07
Lbs. Total Nitrogen Removed per Lb. Dry Solids Wasted	0.066	0.129	0.066	0.060	0.453

TABLE 36.

COMPARATIVE SUMMARY OF AIR AND OXYGENATION TREATMENT SYSTEMS PERFORMANCE

	Phase I Operation		Phase II Operation	Phase III Operation	
	Air System	Oxygenation System	Oxygenation System	Air System	Oxygenation System
Ft^3 Air Utilized/gal. Sewage Treated	2.89	-	-	4.32	-
Feed Oxygen Flow Rate (cfm,NTP)	-	18.3	31.0	-	19.0
Gas Exhaust Rate from Oxygen System Final Stage (cfm,NTP)	-	1.74	4.04	-	3.22
Oxygen System Exhaust Gas Composition as % Oxygen	-	46	55	-	51
Overall % of Feed Oxygen Utilized	-	95.5	92.7	-	91.4
Lbs. Oxygen Utilized/Lb. BOD Consumed	-	0.94	0.96	-	0.72
Total Power Required for Liquid Mixing (HP)	-	26.0	11.1	-	11.7
Total Power Required for Gas Recirculation (HP)	-	2.6	9.2	-	10.6
Overall Power Utilized as HP per 1000 Gal. Mixed-liquor	-	0.088	0.125	-	0.137
Total Air Blower Power Utilized (HP)	141	-	-	138.5	-

In Phase II of the evaluation of the oxygenation unit, one quarter of the available aeration tank capacity was utilized to treat the entire wastewater flow influent to the plant. Under these test conditions, virtually all operating parameters differed significantly from conventional air aeration practice. As shown in Table 33, aeration detention times were quite low (1.2 hours) while mixed-liquor suspended solids concentrations (MLSS - 6980 mg/l and MLVSS - 4450 mg/l) were higher than practically feasible in an air aerated system. As a result of the combination of staging of the high oxygen partial pressure aerating gas and highly efficient gas-liquid contacting devices, a mixed-liquor D.O. concentration (9.0 mg/l), unattainable with air as the aerating gas, was maintained consistently. As was the case in Phase I operation, a highly flocculant and readily settleable mixed-liquor biomass was formed which permitted an average recycle sludge solids concentration of nearly 3% (29,560 mg/l). When manual operation of the clarifier sludge return pumps (air lifts) was closely monitored, a recycle solids concentration of 4.5% was achieved with no apparent undesirable effects to the biomass. As a result of the settling characteristics of the activated sludge, the recycle sludge flow rate as a percent of wastewater flow was not greater than that employed in conventional air aerated systems even though the MLSS level was about three times higher. The food/biomass ratio for this period of operation averaged 0.790 lbs. BOD/day/lb. MLVSS. On an aeration tank volume basis, average organic loadings of 212.5 lbs. BOD/day/1000 ft^3 of mixed-liquor were applied. In general, this combination of factors characteristic of high rate treatment with oxygenation are not practically or economically achievable with conventional air aeration.

In Phase III of the evaluation program,the summary data in Table 33 and 34 indicate that the oxygenation unit was operated under conditions intermediate to the extremes noted in Phase I and II. The three stage gas-liquid contacting system was operated at an average MLSS concentration of 6190 mg/l and mixed-liquor dissolved oxygen concentrations of 8.0 mg/l. The food/biomass ratio was 0.554 lbs. BOD/day/lb. MLVSS and the organic loading was 144.8 lbs. BOD/day/1000 ft^3 mixed-liquor. The oxygenation system high sludge recycle rate, compared to Phase I and II, and commensurately lower recycle sludge concentration have been shown earlier to result from hydraulic limitations of the sludge recycle system which prevented operation in the preferred mode. In general, the operation of the oxygenation system in Phase III was also conducted under conditions not practically feasible for air aeration. Treatment rates were, however, reduced from those of Phase II since only half of the influent wastewater was treated. The resulting average aeration detention time was approximately 2.0 hours (inflow + recycle sludge flow) for the test period.

The air aeration system, operated parallel to the oxygenation unit in Phase III, was controlled somewhat differently from conventional practice. It was the intent of this comparison to evaluate the oxygenation system as described in relation to a high rate air aerated system with comparable liquid staging. Use of staging baffles in only one-half the aeration tank volume employed in Phase I, permitted operation of an air system having three liquid stages at an aeration detention time of 2.6 hours. MLVSS concentrations were increased to accommodate a higher removal rate and an uneconomically high aeration rate (4.32 ft^3/gal. waste treated) was used to maintain dissolved oxygen concentrations in the mixed-liquor. Under conditions of a resulting food/biomass ratio of 0.841 lbs. BOD/day/lb. MLVSS and an organic loading of 128.9 lbs. BOD/day/1000 ft^3 of mixed-liquor, it was not generally possible to maintain mixed-liquor D.O. concentrations above oxygen limiting conditions at MLVSS concentration of > 2600 mg/l.

The data summarized in Tables 34 and 35 illustrate the overall treatment performance of the oxygenation system under the wide variety of operating conditions evaluated. Removals of BOD, COD, and influent suspended solids (TSS) for the oxygenation systems clearly surpass those obtained in the parallel air aeration system. During two different phases (II and III) of the evaluation, operating conditions and treatment rates for the oxygenation system were such that similar operation of conventional air aerated systems would not be economically or practically feasible. Where a parallel comparison of oxygenation system performance with the air aeration system was possible (Phases I and III), it was necessary to use uneconomically high aeration rates for the air system to approach the performance level of the oxygenation unit. Examination of the weekly average performance data shown in the Results reveals that treatment performance by the oxygenation unit was consistently high. Removals of BOD, COD and TSS during the day, the highest organic and hydraulic loading period, did not differ importantly from removals during the night, the period of lowest hydraulic and organic loading.

Total phosphate removals by the air and oxygenation treatment systems did not differ to any great extent where direct comparisons were made. The phosphate removals shown for Phases I and III of the evaluation are relatively high for both systems in comparison to those removal values generally reported for biological secondary treatment. A possible explanation for this high phosphate removal is the hardness and alkalinity of the municipal water supply at Batavia, New York. The U. S. Geological Survey has reported an average alkalinity of 185 mg/l as $CaCO_3$ and an average hardness of 225 mg/l as $CaCO_3$ for finished water used by the municipality. The total phosphate removal listed for Phase II of the evaluation may not be valid due to the admission of anaerobic digestor supernatant to the system in a manner which prevented sampling with the influent wastewater.

Removals of ammonia and total Kjeldahl nitrogen by the air and oxygenation systems were equivalent where these comparisons were made in Phases I and III operation. Overall nitrogen removals for each system were approximately 36%. The level of nitrification in neither the air nor oxygenation system was remarkable. Both systems achieved on the average a 7% degree of nitrification based on the fraction of total effluent nitrogen as nitrate in various evaluation periods. These values did not exceed 10% of the effluent ammonia nitrogen content. Essentially no nitrification was accomplished in the air system during the Phase I evaluation period.

It would seem reasonable to have expected significant nitrification by the oxygenation system in particular due to the consistently high dissolved oxygen level maintained in the mixed-liquor throughout the various test periods. The reason for lack of extensive nitrification in view of this consistently high oxygen level and relatively long apparent sludge age was not quantitatively ascertained. The relatively low temperature of the feed wastewater (see Table 33) may possibly be implicated, but such speculation is not factually based. Operating criteria essential to establishing nitrification in an oxygenation system will be one of the objectives of future studies.

Further data with respect to nitrogen removals by the air and oxygenation systems are summarized and compared in Table 35 where average lbs. total nitrogen removed/lb. dry solids wasted are listed for each phase of operation. As indicated, the air system activated sludge contained an average of 6.0-6.6% nitrogen (on a dry weight basis) for Phases I and III test periods. In Phases I and II operation, the waste activated sludges from the oxygenation system contained an average of 12.9% and 6.6% nitrogen, respectively. These values are all within the range of 4%-14% nitrogen per dry weight unit of waste activated sludge most frequently reported in the literature for domestic waste treatment. The Phase III oxygenation system value of 45.3% of nitrogen per dry weight unit of waste activated sludge, however, is definitely not an expected or acceptable value.

Possible reasons for this discrepancy are many, but in the present study may be reasonably limited to two most probable ones, inaccurate measurement of waste sludge volume or nitrogen loss from the system to the atmosphere by denitrification. The former reason appears least probable of the two since oxygen system waste activated sludge volume was recorded on two different magnetic flow meters in series with the totalizer results being checked for coincidence. The second reason, denitrification, is more probable than the first but cannot be ascertained quantitatively. This question will be resolved in future studies by direct total nitrogen analysis of the waste activated sludge, a determination not carried out in the work reported here.

In all phases of the oxygenation system evaluation, the remarkable physical properties of the activated sludge produced were consistently apparent. The mixed-liquor settled quite rapidly and compacted well to yield evaluation period average SVI values as low as 36 during Phase II operation. Individual SVI values of as low as 20 were reported at various times throughout the three operational periods. The initial settling velocities for the oxygenation system mixed-liquor in Phases II and III are particularly striking in comparison to air system activated sludge values. These velocities are related to the total suspended solids content of the mixed-liquor settling sample. Despite the hindrance imposed by an approximate two-fold higher suspended solids level in the oxygenation system mixed-liquor, initial settling rates were comparable to those of the air system activated sludge. The rapidly settling sludge from the oxygenation system consistently gives a descrete liquid solids interface during the settling test. These factors, rapid settling and low SVI values, are the basis for the relatively high recycle sludge solids concentration observed in all phases of oxygenation system evaluation. In Phase II operation,an average 3% recycle sludge suspended solids concentration was possible even with the limitations in clarifier operation previously discussed. In Phase III operation,an average recycle suspended solids concentration of approximately 1.9% was observed for the oxygenation system compared to approximately 1.7% for the air aerated system; however, the sludge recycle rate of the oxygenation system was more than twice that of the air aerated system (the reason for this higher rate was pointed out in the Results). Limited periods of testing in Phase II of the evaluation indicated that an average recycle sludge suspended solids concentration of 4.0-4.5% may be possible on a sustained basis with careful operation of the oxygenation system clarifier and recycle pumps. Additional results of limited vacuum filtration tests during this period showed that waste recycle sludge could be filtered directly (after floculation with lime and $FeCl_3$ conditioner) without additional thickening. The implication of these observations is that by taking advantage of a low recycle rate, a waste activated sludge thickener may not be necessary in treatment facilities using the oxygenation system. Certainly,the demonstrated ability of the system to yield substantially higher recycle sludge suspended solids concentrations and thus sustain high mixed-liquor solids concentrations at low recycle sludge flow rates is one of the key factors which makes it economically possible to operate the high rate treatment process evaluated here. This property of the oxygen aerated activated sludge system previously observed in pilot scale studies and shown in the full scale tests described here has not been previously recognized or reported in the literature by others.

Determination of Cell Yield and Endogenous Respiration Coefficients

Both Eckenfelder and O'Connor (12) and Eckhoff and Jenkins (13) have derived expressions for steady state conditions which may be used to describe the production of excess VSS by an activated sludge treatment system. As an approximation here, the COD based expression of Eckhoff and Jenkins may be used with BOD removal as the basis if one assumes a constant relationship between the rates of BOD and COD removal. A plot derived from the test data obtained in this study is shown in Figure 12 whereby the cell dilution rate as lbs. excess VSS/day/lb. MLVSS is related to the specific reaction rate as lbs. BOD removed/day/lb. MLVSS. An equation of the linear relationship obtained is of the form

$$y = mx + b$$

where m = cell yield coefficient as lbs. VSS produced per lb. BOD removed

and b = endogenous respiration coefficient as day^{-1}

In the present comparison of air and oxygen aeration, it was useful to attempt to obtain comparative values for the cell yield and endogenous respiration coefficients. For such values to be meaningful in the context of the available literature, BOD removals should have been calculated on a soluble BOD basis. This was not possible since whole rather than filtered effluent samples were used for BOD determinations in the work presented here. However, this constraint was common to both systems and the values are meaningful on a comparative basis.

Attempting to plot the data directly, as shown in Figure 12, did not provide useful correlations, probably a result of the extreme unsteady state conditions in both the air and oxygenation systems. As a result, the relationship shown was obtained in an indirect manner. Both the cell dilution rates (lbs. excess VSS/day/lb. MLVSS) and specific reaction velocities (lbs. BOD removed/day/lb. MLVSS) were plotted as a function of a common parameter, aeration detention time (based on raw wastewater plus recycle flow), as shown in Figures 13 and 14. A good correlation of specific

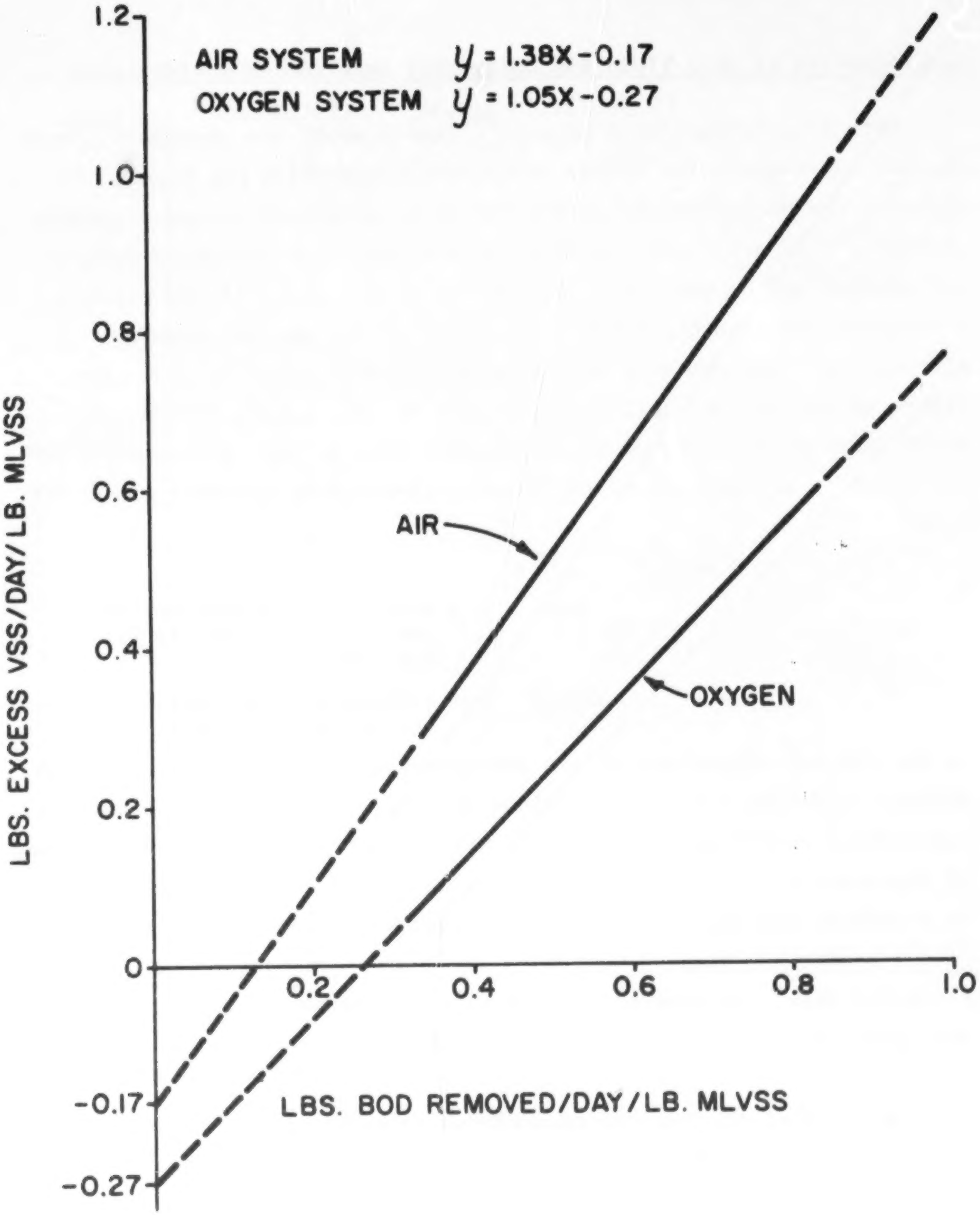


FIGURE 12

CORRELATION OF EXCESS VSS FROM AIR AND OXYGEN SYSTEMS WITH BOD REMOVAL AND LBS. MLVSS

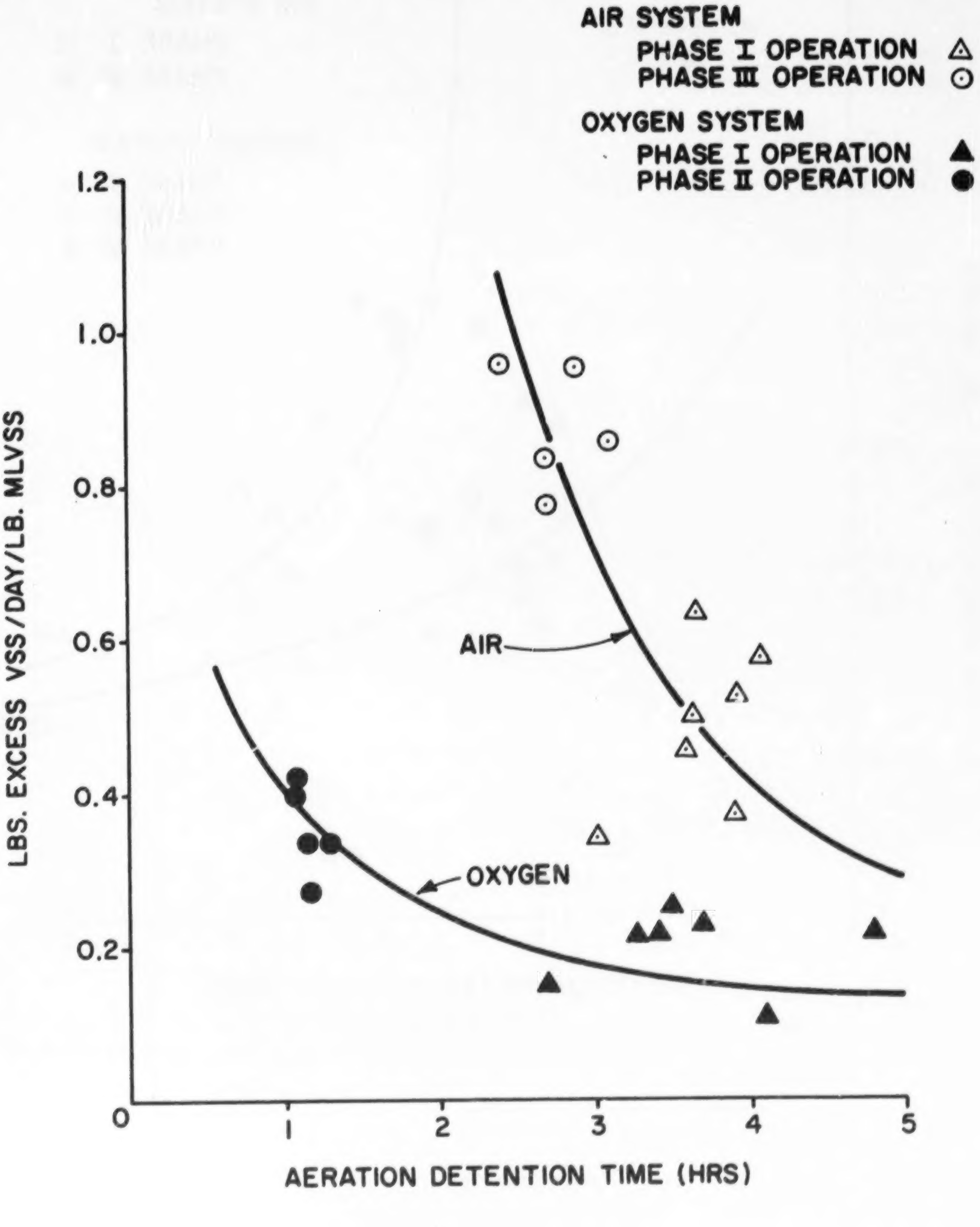


FIGURE 13

COMPARATIVE CORRELATION OF EXCESS VSS
AND MLVSS WITH AERATION DETENTION TIME

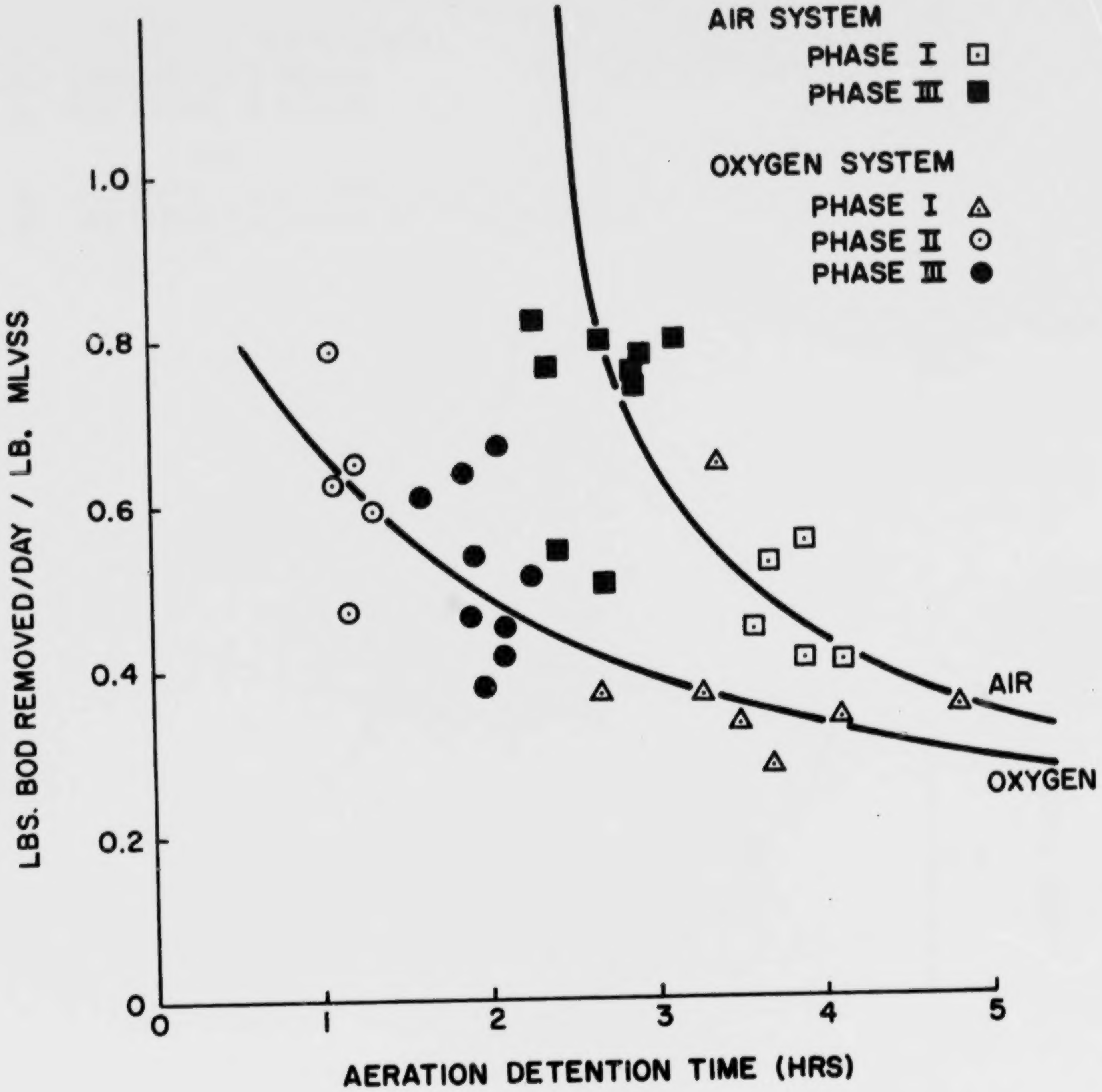


FIGURE 14

COMPARATIVE CORRELATION OF LBS. BOD REMOVED AND MLVSS WITH AERATION DETENTION TIME reaction velocity was obtained as shown in Figure 14 for both the air and oxygenation systems. Similarly, the correlation of cell dilution rate to aeration detention time is shown in Figure 13. For probable reasons to be subsequently discussed, the cell dilution rates for Phase III operating data from the oxygenation system did not correlate and were, by necessity, excluded. Using values obtained for comparable aeration detention times from the lines drawn in Figures 13 and 14, it was possible to construct the relationship shown in Figure 12 relating cell dilution rates to specific reaction velocities with respect to BOD. The equations of the two lines constructed in Figure 12 are shown where

for the air system $$\frac{\text{lbs. Excess VSS}}{\text{Day X lbs. MLVSS}} = 1.38\,\frac{\text{lbs. BOD Removed}}{\text{Day X lbs. MLVSS}} - 0.17\ \text{day}^{-1}$$

and in the oxygenation system $$\frac{\text{lbs. Excess VSS}}{\text{Day X lbs. MLVSS}} = 1.05\,\frac{\text{Lbs. BOD Removed}}{\text{Day X lbs. MLVSS}} - 0.27\ \text{day}^{-1}$$

The cell yield coefficients obtained in each case clearly indicate by comparison relative differences in excess volatile solids production which are expected when one views the sludge production values shown in Table 34 for the air and oxygenation treatment systems. The air system produced substantially more VSS per day than did the oxygenation system, particularly at the relatively high rate treatment conditions in Phase III. Also, the apparent endogenous respiration coefficients shown here are nearly two-fold greater in the case of the oxygenation system activated sludge than in the case of the air aeration system.

The apparent yield coefficients and endogenous respiration coefficients obtained for both the air and oxygen systems appear consistent with the results in all but one instance, Phase III oxygenation system operation. In this period of the evaluation program, approximately seven-fold more excess volatile solids were produced by the air system than by the oxygenation system even though both systems were operating at nearly the same volumetric organic loadings. It was for this reason, and the previously discussed discrepancy in the nitrogen balance of the oxygenation system, that the Phase III oxygen system cell dilution rate data was not used in the correlation shown in Figure 12. The yield and endogenous respiration coefficients shown from Figure 12 do not reflect, therefore, the extremely low excess VSS production by the oxygenation system during Phase III operation.

It is clear from the comparative excess sludge wastage values in the Results as well as the kinetic evaluation of these data in the foregoing that the oxygenation system is characterized by the production of substantially less volatile solids than the parallel air system. The reasons for this difference may be quite complex and will require further investigation. One factor of note is the high food/biomass ratio apparent in the air system, particularly in Phase III operation,coupled with periodic dissolved oxygen limiting conditions. In this instance,a substantial portion of BOD may have been wasted from the system in unoxidized form after immeshment in the bacterial floc (an observation supported by the odor and appearance of recycle sludge or waste sludge). This would tend to increase the apparent cell dilution rate and specific reaction velocity over that obtained where oxygen is not limiting. In Figure 12,the results of this circumstance would be an increase in the slope of the air system correlation, a higher yield coefficient, and an increase in the y-axis intercept, the endogenous respiration coefficient. If this condition were so, evaluation of the air system operating conditions between the extremes of Phase I and III might reveal the relationship in Figure 12 to be biphasic. Obviously,this explanation cannot be tested by the results presented; however, the synthesis of several observations strongly suggest this possible reason for the apparent high yield and endogenous respiration coefficients for the air aeration system.

Another factor of probable importance in reduction of excess volatile solids production by the oxygenation system is the sustained high MLVSS concentrations, particularly in Phase II and III, and high dissolved oxygen concentration. While organic loadings were generally high for the oxygenation system, the ability to operate at very high MLVSS concentrations yielded food/biomass ratios generally lower than the air aeration system in both Phases I and III of the evaluation (see Table 34). Examination of diurnal variations in food/biomass ratios in the week by week values in the Results reveal 3-4 fold differences from day to night. This observation coupled with increased oxygenation detention times at night indicates the possibility of extensive auto-oxidation of the biomass at night following an extensive period of synthesis during the day. Partial confirmation of this occurrence may be observed in the following sections of the Discussion where day to night variations in oxygen utilization/lb. BOD removed are shown. If auto-oxidation of biomass VSS does occur extensively in the various phases of oxygenation system evaluation, it must result from a daily period of food limiting conditions. This would suggest the sludge reduction due to endogenous respiration may in fact be greater than indicated in Figure 12.

A further evaluation of oxygenation system data was done to obtain values for respiration rates associated with BOD removal and endogenous activity of the cell mass. According to Eckenfelder and O'Conner(14), these values may be obtained from the slope and intercept of a plot correlating lbs. oxygen required and lbs. MLVSS with lbs. BOD removed and lbs. MLVSS. Using oxygenation system data, this relationship is shown in Figure 15 and reveals that the system requires 0.530 lbs. oxygen/lb. BOD removed for new cell synthesis and 0.166 lbs. oxygen per hour to sustain the endogenous respiration for each pound of MLVSS under aeration.

Attempts to kinetically evaluate the parallel air and oxygen treatment systems using the same set of assumptions in each case have posed more questions than can be answered by this work. For example, the means by which substantially less inert suspended solids are wasted from the oxygenation system, compared to the air system, even though the same quantity of inert solids suspended is contained in the common feed wastewater to the two systems is not known at this time. One possible explanation is that some fraction of the inert suspended solids entering the oxygenation system are leaving as inert dissolved solids in the final effluent. This can not be verified, however, as dissolved solids balances were not carried out for either system. In part, the unanswered questions are due to the necessarily limited number of operating conditions evaluated and the experimental program having been formulated to establish feasibility of oxygenation and its relative economics. One major factor

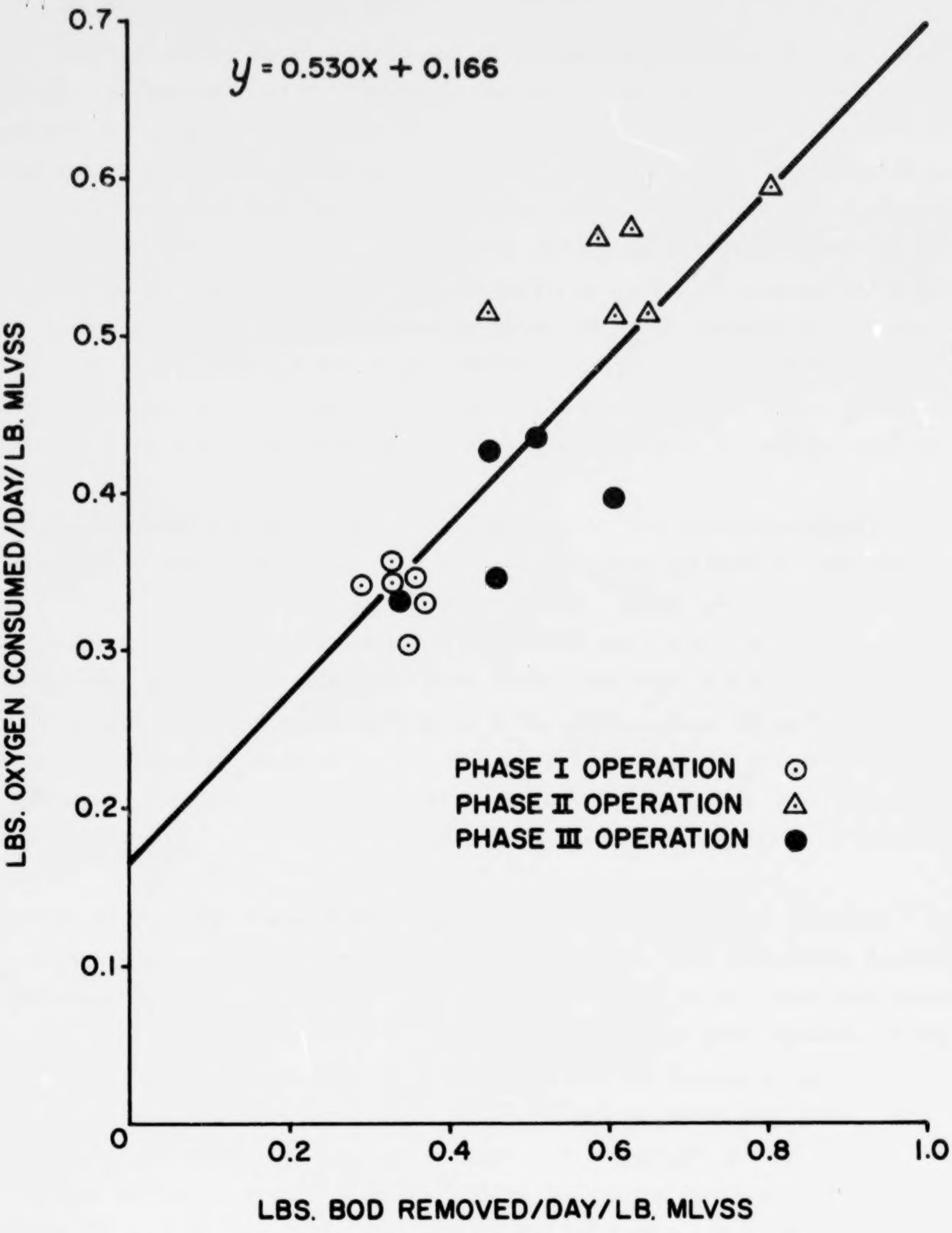


FIGURE 15

OXYGENATION SYSTEM CORRELATION OF LBS. OXYGEN CONSUMED/LB. MLVSS WITH LBS. BOD REMOVED/LB. MLVSS which should not be overlooked in the comparison, however, is the character of the biomass produced by the oxygenation unit. Previous laboratory and pilot plant studies have indicated that activated sludge acclimated to high dissolved oxygen concentrations with high aerating gas oxygen partial pressures may differ distinctly (microbiologically) from activated sludge developed with air aeration at relatively low dissolved oxygen concentrations. The ramifications of this observation, though beyond the scope of the present discussion, are possibly a key factor in performance differences noted between the air and oxygenation systems in this work. It is possible that the same set of assumptions cannot be applied in describing the two systems kinetically.

Relationship of Oxygen Utilized and BOD Consumed to Food/Biomass Ratio and Oxygenation Detention Time

It has been shown in the Results and Table 33 that substantially less volatile solids were produced by the oxygenation treatment system per lb. of BOD applied than by the air aeration system operated under nearly identical organic loading conditions in test Phases I and III. As pointed out in the foregoing Discussion section, a higher apparent rate of microbial metabolism and lower net rate of excess sludge production due to auto-oxidation of the oxygenation system biomass could explain these differences. In Figure 16 are empirical correlations of lbs. oxygen utilized/lb. BOD removed to food/biomass ratio per hour aeration detention time (includes raw waste + recycle flow). Detention time is arbitrarily included as a parameter to permit each period of operation to be distinguished from the others. Night and daytime weekly average values are shown in each instance. The correlation emphasizes the 3 to 4 fold apparent increase in the amount of oxygen required to remove equivalent weight units of BOD at night as compared to daytime. Also emphasized is the difference in food/biomass ratio from day to night for a given phase of operation. Aeration detention time differences from day to night have little effect in the observed spread in food/biomass ratio since detention times are fairly constant for the day and night periods in a given phase of the test program.

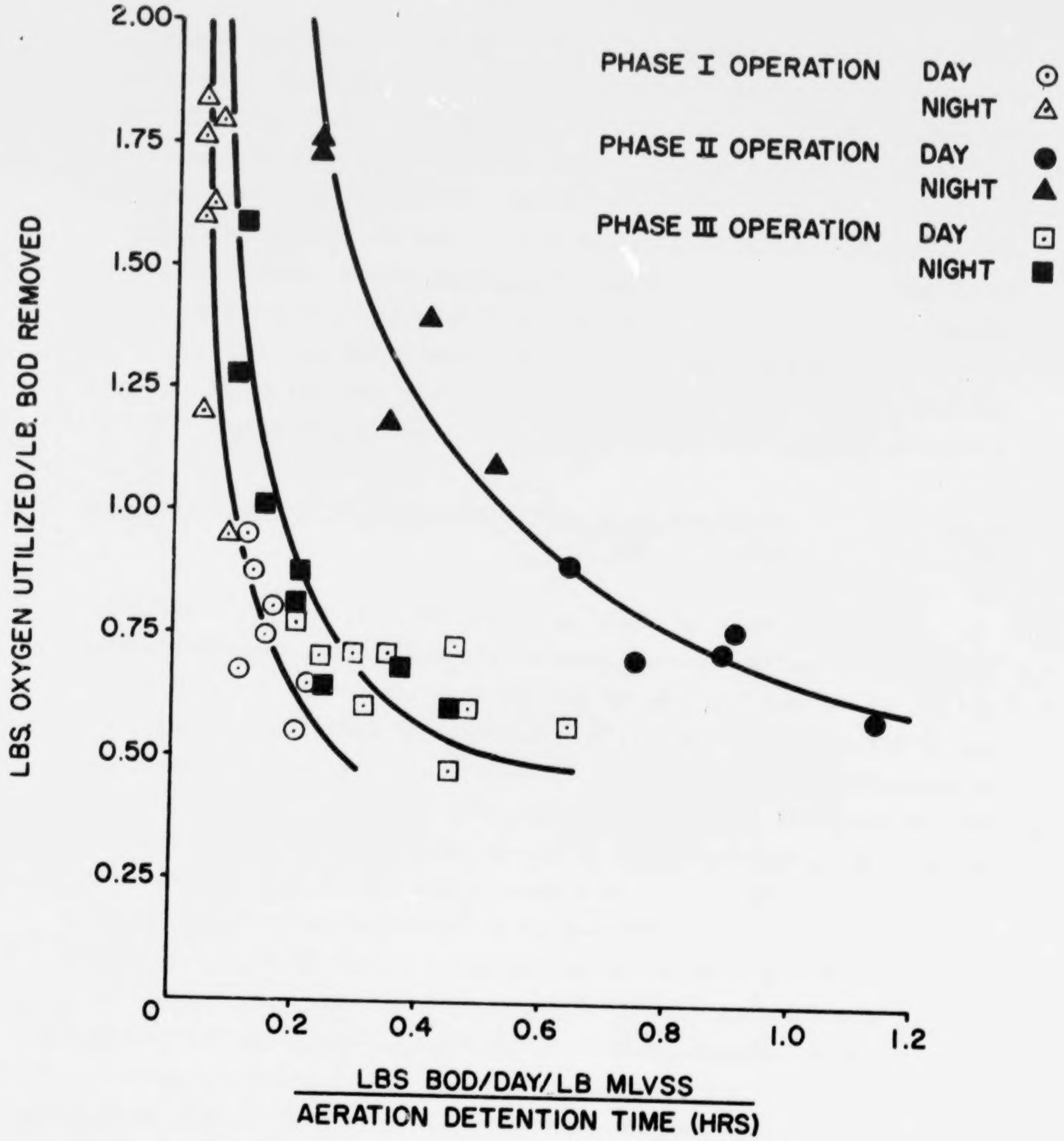


FIGURE 16

OXYGENATION SYSTEM CORRELATIONS OF OXYGEN UTILIZATION/LB BOD REMOVED WITH FOOD/BIOMASS AND AERATION DETENTION TIME

Literature sources in general indicate approximately 0.75 lbs. of oxygen as necessary for the metabolism of a pound of BOD in domestic waste treatment using conventional air aeration techniques. This observation is consistent with the oxygen utilization rate during the daytime periods of operation in Figure 16 when food/biomass ratios are highest and aeration detention times are lower relative to night time operation. During night time operation when food/biomass ratios were one-fourth as high as daytime, and MLVSS concentrations were approximately the same, the oxygen utilization rate per lb. BOD removed was 2.0-2.5 times greater.

The implication of the foregoing observation may be two-fold. First, it appears that the oxygenation system biomass at high food loadings during the daytime operation consumed oxygen and BOD for new cell synthesis through a mechanism not unlike conventional systems even though the apparent treatment rate was higher in some instances. Secondly, and most importantly, the oxygen utilization as an apparent function of BOD removed at night (when food/biomass ratios are low) would indicate significant endogenous respiration or auto-oxidation of the volatile solids yield from daytime synthesis. The overall result of this "partial stabilization process" would be a significant reduction in overall excess solids production by the oxygenation system, a condition revealed in the Results. On a comparative basis, this reduction would be all the more striking where the air system operated in parallel was subjected to unusually high organic loadings, resulting in a decrease in lbs. oxygen available for consumption of BOD.

The foregoing proposed explanation is consistent with the results contained in this report and most probably is one of the major factors in the difference in solids production by the two systems. The conclusion that it is the only factor involved is not warranted by these results. A practical conclusion warranted by these results, however, is that in all oxygenation system evaluation phases a substantial reduction in the amount of sludge produced is apparent in direct comparison to the air aerated activated sludge system. Therefore, one of the recognized advantages of the oxygenation system in domestic waste treatment is a potential reduction in the amount of waste sludge disposed and thus in the capital investment for and operating cost of this phase of treatment plant operation.

Relationship of BOD Removal to Detention Time and MLVSS Concentrations

For purposes of the economic comparison of conventional air aerated and oxygenation systems, it was convenient to develop a correlation relating BOD removal to treatment detention time and mixed-liquor volatile solids concentrations which would be consistent for both systems. Thereby, a means of developing comparative cost data for equivalent treatment efficiencies with respect to BOD removal would be provided. The desired relationship was obtained and is shown in Figure 17. In this plot, weekly average data for performance of both the air and oxygenation systems is presented. Percent BOD removed/hour actual treatment detention time/MLVSS concentration plotted against hours actual treatment detention time x MLVSS concentration (actual detention time considers both raw waste and recycle sludge flow) yields a line which is described by the indicated equation where

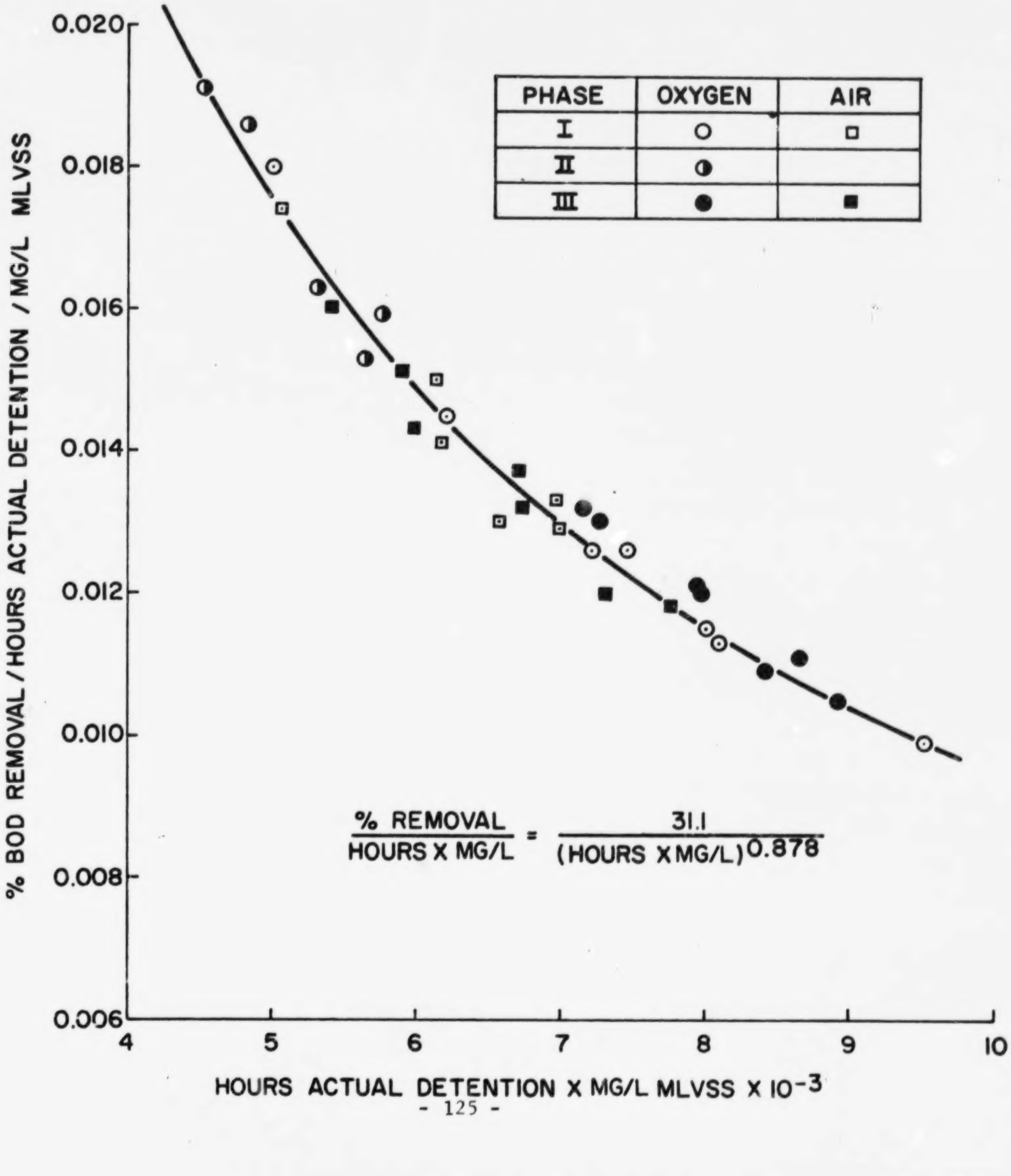
$$\frac{\%\ \text{BOD Removal}}{\text{Hours x mg/l}} = \frac{31.1}{(\text{Hours x mg/l})^{0.878}}$$

$$\%\ \text{BOD Removal} = (31.1)\ (\text{Hours Actual Detention Time x mg/l MLVSS})^{0.122}$$

This empirical treatment of the data provides a convenient means of describing the probable combinations of system operating conditions necessary to achieve equivalent treatment efficiencies (BOD removals) in both the conventional air and oxygenation systems, providing that oxygen concentrations are not seriously limiting in either case. A plot of oxygenation system data alone yields a straight line relationship. The curvilinear form of the combined data relationship shown, attributed to air system values, probably is a result of oxygen limiting conditions in this system, particularly in Phase II of the evaluation. Also, it must be remembered that the air

CORRELATION OF % BOD REMOVAL
AT BATAVIA WITH ACTUAL DETENTION
PERIOD AND MG/L MLVSS

FIGURE 17



system could not be operated at MLVSS concentrations above 2600 mg/l in the Phase III evaluation without serious oxygen limitation while operation of the oxygenation system was possible at nearly two times this MLVSS concentration. Actual operational limitations then existed for the air system which were not present for the oxygenation unit. Despite these considerations, the relationship in Figure 17 provides a valuable, although approximate tool, for relating the performance of the two systems in this work.

AIR AND OXYGENATION SYSTEMS

ECONOMIC COMPARISON

Introduction

The Batavia test data presented here have demonstrated the practical feasibility of economically operating a high rate, high solids activated sludge system utilizing oxygen aeration to maintain high dissolved oxygen levels. The data have substantiated that high treatment quality can be routinely attained under operating conditions that heretofore would have been considered impractical and uneconomical.

Observations made have eluded to the potential for cost reduction and treatment quality improvement inherent in the use of oxygen aeration. The magnitude of these cost reductions will now be quantitatively evaluated in detail. This will be accomplished by consideration of the secondary treatment, sludge disposal, and total plant costs for air and oxygen aerated activated sludge processes treating municipal wastewater for a variety of design conditions over a range of plant sizes varying from 1 to 100 MGD. Specifically, new treatment plant facilities will be costed for each process at four different aeration residence times (and hence overall BOD removals) for plants of 1, 6, 30, and 100 MGD capacity. It must be realized, however, that it is practically infeasible to evaluate all potential modifications and ramifications of each of the two processes. Nevertheless, it is felt that the following comparison is fairly respresentative of the relative cost effectiveness of the two processes for the treatment of typical municipal wastewater. It should be pointed out, however, that specific unusual situations could significantly alter the conclusions in either direction relative to the costs presented here. Of particular note in this repect is the applicability of the oxygenation process in upgrading existing activated sludge plants to either increase capacity and/or improve treatment performance. In these cases the advantage of the oxygenation process relative to other alternatives will be significantly greater than in the case of comparative new plant construction costs presented here.

Of particular note in these comparisons will be the mode of oxygen supply for the oxygenation systems. An on-site oxygen gas generator will be costed as an integral part of the overall oxygenation system. Hence, the cost of supplying oxygen will be evaluated on an amortized capital investment and operating cost basis just as any other piece of equipment in the overall wastewater treatment facility. Alternative oxygen supply methods would include: 1) transport of liquid oxygen to the waste treatment plant for on-site evaporation, and 2) use of a gaseous pipeline supply if available in the plant vicinity. Separate costs are provided herein for on-site gas generation plants in order that cost comparisons can be made for alternative supply methods.

Oxygen generation plants are commercially available from the Linde Division of Union Carbide Corporation and others. Oxygen gas generation plants are of a completely automatic type which run unattended except for infrequent inspection by plant operators.

Basis of Economic Comparison

Bioreactor Design

The relative cost effectiveness of oxygen and air aerated secondary activated sludge systems were evaluated by considering a range of practical process design conditions required for each system to accomplish comparable overall removal levels of BOD and suspended solids. The range of the primary process design parameters evaluated are summarized in Table 37. A raw wastewater BOD of 200 mg/l is assumed in all cases with a 35 percent removal in the primary clarifiers yielding a primary effluent BOD of 130 mg/l influent to the activated sludge aeration tanks.

TABLE 37

BASIC PROCESS DESIGN DATA FOR AIR AND OXYGENATION SYSTEMS ECONOMIC COMPARISON

	Oxygen System	Air System
Raw Wastewater BOD, mg/l	200	200
Primary Effluent, BOD, mg/l	130	130
Aeration Detention Time,(hr.) (based on wastewater flow only)	1.0,1.33,1.66,2.0	3.0,4.0,5.0,6.0
MLSS Concentration, mg/l	6800	2500
MLVSS/MLSS	0.65	0.65
MLVSS Concentration, mg/l	4420	1625
Recycle Sludge Concentration,mg/l	40,000	10,000
Mixed-Liquor D.O., mg/l	8.0	1.0
Aeration Rate (cf NTP)/gallon)	—	0.8, 1.6, 2.4
Oxygenation System Absorption Efficiency (%)	90-92	—

The relative economics of secondary treatment with and without primary clarifiers are not considered a a part of this study. The results reported here show that raw, prescreened and comminuted sewage can be directly treated in the secondary aeration tanks with the attainment of high treatment quality in relatively short aeration detention times. Due to the predominance of primary clarifiers in new plant design, however, their use is assumed throughout these economic comparisons. Further consideration of this parameter will be undertaken in future work.

The design MLSS concentrations assumed are 2500 mg/l and 6800 mg/l for the air and oxygenation systems respectively with each system having an MLVSS/MLSS ratio of 0.65. This corresponds to MLVSS concentrations of 1625 mg/l and 4420 mg/l in each case. The MLSS are assumed to concentrate to 10,000 mg/l and 40,000 mg/l in the final clarifier underflow, respectively, for the air and oxygenation systems.

Four different aeration residence times (based on raw wastewater flow only) were evaluated for each system with the constant MLSS and MLVSS concentration values stated. This corresponds to a range of BOD removals in the secondary system of approximately 84% to 92% for each system as summarized in Table 38. The volumetric organic loadings and the food/biomass ratios for each of the system designs are summarized in Table 39. As shown,the oxygenation system is designed to operate at volumetric organic loadings of from 97 to 194 lbs. BOD/day/1000 ft^3 of mixed-liquor in contrast to values of 32 to 65 for the air system. The food/biomass ratios, however, are comparable due to the much higher MLVSS concentration that can be practically and economically maintained in the oxygenation system.

The basis for these bioreactor design specifications and relative performance data predictions are based on data obtained during this study and are shown in Figure 17 wherein the percent BOD removal and MLVSS concentration is plotted as a function of the aeration residence time and MLVSS concentrations for both the air and oxygenation treatment systems. These data have already been discussed. A few comments are in order,

TABLE 38

SUMMARY OF AERATION RESIDENCE TIMES AND BOD REMOVALS FOR AIR AND OXYGENATION SYSTEM ECONOMIC COMPARISONS

Air System Aeration Detention Time-Hrs. (based on Raw Flow)	Percent BOD Removal Secondary System	Oxygen System Aeration Detention Time-Hrs. (based on Raw Flow)
3.0	84	1.00
4.0	87	1.33
5.0	90	1.66
6.0	92	2.00

TABLE 39

SUMMARY OF VOLUMETRIC ORGANIC LOADINGS AND FOOD TO BIOMASS RATIOS FOR AIR AND OXYGENATION SYSTEM DESIGNS

Aeration Detention Time-Hrs(based on raw flow)	MLVSS Conc. mg/l	Volumetric Organic Loading (lbs BOD/Day 1000 ft^3)	Food to Biomass Ratio (lb BOD/Day lb MLVSS)
	OXYGENATION SYSTEM		
1.00	4420	194	.70
1.33	4420	146	.53
1.66	4420	117	.42
2.00	4420	97	.35
	AIR SYSTEM		
3.0	1625	65	.64
4.0	1625	49	.48
5.0	1625	39	.38
6.0	1625	32	.32

however, regarding the data in Tables 38 and 39, and the effect of the specific reactor design flow configuration upon the performance data shown in these tables. By reactor design flow configuration, the commonly used modifications of the activated sludge process are referred to. These are the conventional plug flow, step aeration, and completely mixed reactors. This range of process modification includes the complete spectrum of reactor hydraulic design configuration from that of a completely non-mixed (plug flow type) to that of a completely back-mixed, single stage reactor system having uniform composition throughout. Certainly these factors have a significant influence on overall system performance and hence reactor volume to achieve a specified level of overall BOD removal at a given MLVSS concentration. As is well known, they also significantly influence the aeration system performance requirements to sustain an adequate dissolved oxygen level. It is beyond the scope of this study, however, to evaluate the effect of these various reactor configurations upon the required system volume. From an overall economic standpoint, the only factor of any relative significance is the required aeration tank volume itself. Within the limits of the three to six hour aeration detention times, shown in Tables 38 and 39 for the air aeration systems, it is not felt that these various reactor flow modifications are of significant influence provided that oxygen transfer limitations are not a factor. Data for completely mixed and step aeration systems show comparable performance to that of plug flow systems within this range of solids levels and aeration detention times. Current practice seems to indicate a preference for step aeration design at an aeration residence time of from three to four hours to accomplish about 90 percent BOD removal. From an economic standpoint these designs would be comparable to those projected here even though the comparative design herein is based on conventional plug flow operation.

Aeration and Oxygenation System Design

The oxygen requirements for the process designs used were determined from the respective synthsis and endogenous respiration coefficients discussed earlier. These coefficients are 0.530 lbs. oxygen/lb. of BOD removed and 0.166 lbs. oxygen/day per pound of VSS under aeration. The respective oxygen requirements for the two systems are contained in Table 40 for the various process design conditions outlined. As shown, the oxygen requirements are somewhat higher for the oxygen system due to the higher volatile solids levels under aeration and the higher rate of oxidation of the volatile solids even though the aeration detention times are lower.

Certainly many different types of air aeration systems are in use today in activated sludge plants. Due to the predominence of diffused air systems in municipal wastewater treatment facilities, however, this type of air aeration system was chosen as the basis for the economic comparison. The capital costs and power requirements of these systems are documented here in sufficient detail so that other types of aeration equipment could be compared if desired. Considerable care must be exercised, however, in evaluating capital, operating and maintenance, and power requirements of alternative aeration systems. Performance must be evaluated under the actual mixed-liquor aeration conditions experienced in the activated sludge process which are often quite different from those normally used in the rating of mechanical aerators.

Furthermore, since the efficiency of diffused air systems as reported in the literature is widely variant and often confusing regarding the various claims that are made, a range of air rates and hence oxygen absorption efficiencies were evaluated in order to ascertain the effect of this parameter on the relative economic comparison. For this purpose, air rates of 750, 1500, and 2250 cf(NTP) of air per pound of BOD applied were evaluated. These air flow rates are equivalent to aeration rates of 0.8, 1.6, and 2.4 cf(NTP) of air per gallon of sewage for the assumed process conditions outlined. The air rate of 1500 cf(NTP) per pound of BOD applied or 1.6 cf(NTP) of air per gallon of sewage would correspond to the commonly accepted Ten States Standards. Because many claims are made regarding diffusers of significantly higher efficiency, however, an air rate of half this amount or 0.8 cf(NTP) per gallon was evaluated.

TABLE 40

COMPARATIVE OXYGEN SUPPLY AND CONSUMPTION RATES

Oxygen Supplied and Consumed in Oxygenation System

*Hours Detention	$\frac{\text{Tons } O_2 \text{ Supplied}}{\text{Day x MGD}}$	$\frac{\text{Tons } O_2 \text{ Consumed}}{\text{Day x MGD}}$
1.00	0.46	0.41
1.33	0.51	0.46
1.66	0.57	0.51
2.00	0.61	0.56

	$\frac{\text{Lb. } O_2 \text{ Supplied}}{\text{Lb. BOD Removed}}$	$\frac{\text{Lb. } O_2 \text{ Consumed}}{\text{Lb. BOD Removed}}$
1.00	1.00	0.90
1.33	1.08	0.97
1.66	1.16	1.05
2.00	1.22	1.12

Oxygen Consumed in Air Aeration System

*Hours Detention	$\frac{\text{Tons } O_2 \text{ Consumed}}{\text{Day x MGD}}$	$\frac{\text{Lb. } O_2 \text{ Consumed}}{\text{Lb. BOD Removed}}$
3	0.39	0.85
4	0.44	0.94
5	0.50	1.02
6	0.55	1.11

* Based on Raw Sewage Flow

The blower HP requirements were based on a blower discharge pressure of 8.1 psig and overall electric power requirements were based on a compressor efficiency of 70 percent and a motor efficiency of 95 percent. Similarly, the oxygenation system power requirements were based on compressor and motor efficiencies of 74 and 95 percent for the gas recirculation compressors and agitation units.

The power requirements for the two systems are summarized in Tables 41 and 42 for the various process design conditions. The power requirements for the diffused air system are a function of the assumed air flow rate, whereas the power requirements of the oxygenation system are a function of the aeration residence time. This is, of course, a result of the fact that a considerable portion of the energy is expended in liquid mixing as opposed to gas compression. Consequently, the oxygenation system power requirements are a function of the quantity of mixed-liquor under aeration in addition to the oxygen requirements for BOD removal. The oxygenation systems in all cases are designed for an overall utilization efficiency of 90-92 percent in the secondary aeration tanks.

The data in Tables 41 and 42 show that a substantial overall power savings can be realized through the use of the oxygenation system under all conditions except the extremes of the lowest air rate (0.8 cf(NTP) per gallon) and the longest residence time for the oxygen system (2 hours). Under the above conditions power savings of from 15 to 25 percent can be obtained in the larger plant sizes. At an aeration time of one hour, power savings of from 33 to 45 percent can be obtained compared to even the lowest air aeration rate of 0.8 cf(NTP) per gallon. Compared to an air rate of 1.6 cf(NTP) per gallon, power savings of from 65 to 70 percent can be obtained with the oxygenation system. In all of these comparisons it should also be remembered that the oxygenation systems are operating at dissolved oxygen levels of 8 to 10 mg/l compared to the nominal air aeration dissolved oxygen levels of only 1 to 2 mg/l.

TABLE 41

OXYGENATION SYSTEM POWER REQUIREMENTS

MGD Treated	*Hours Detention	Oxygen Dissolution HP	Oxygen Generation HP	Total Oxygen System HP
1	1.00	6	11	17
	1.33	7	12	19
	1.66	8	14	22
	2.00	10	15	25
6	1.00	37	63	100
	1.33	43	70	113
	1.66	50	78	128
	2.00	57	85	142
30	1.00	180	300	480
	1.33	210	350	560
	1.66	250	385	635
	2.00	280	420	700
100	1.00	510	800	1310
	1.33	620	900	1520
	1.66	740	1000	1740
	2.00	870	1090	1960

* Based on Raw Sewage Flow

TABLE 42

DIFFUSED AIR AERATION POWER REQUIREMENTS

	BLOWER HP REQUIRED		
MGD TREATED	0.8 CF (NTP) OF AIR PER GAL. OF SEWAGE	1.6 CF (NTP) OF AIR PER GAL. OF SEWAGE	2.4 CF (NTP) OF AIR PER GAL. OF SEWAGE
1	25	50	75
6	150	295	440
30	735	1470	2200
100	2450	4880	7350

Primary and Waste Activated Sludge Disposal

As stated earlier, it has been assumed that primary clarifiers are employed in each of the comparative system designs and that these clarifiers remove 35 percent of the influent BOD. From a sludge production standpoint it has been further assumed that the primary clarifiers in each case produce 600 pounds of VSS per day per MGD treated and that this primary sludge has a VSS/TSS ratio of 0.7. With respect to the primary clarifiers and their associated sludge production, therefore, both the air and oxygenation systems are assumed identical.

The foregoing assumption of equivalency is not the case, however, with respect to the waste activated sludge produced and its disposal requirements. As shown in Table 37, the waste activated sludge or recycle sludge from the oxygenation system is assumed to have a concentration of 4 percent suspended solids as compared to that from the air system of 1 percent. This is a consequence of the considerably higher MLSS concentration (6800 mg.1 compared to 2500 mg/l) carried in the oxygenation system and its enhanced settling, compacting and dewatering characteristics. It is further assumed, at 4 percent solids, that the waste activated sludge from the oxygenation system can be disposed of without prior thickening. In contrast, the air system waste activated sludge must be thickened prior to further processing. For purposes of this report, it has been assumed that the waste activated sludge from the air system would be gravity thickened from the 1 percent suspended solids level to 4 percent.

For the 1 MGD plant size, it was assumed that sludge disposal for the combined primary and waste activated sludges would consist of anaerobic digestion and sludge drying beds. Anaerobic digestion, vacuum filtration, and cake disposal by landfill was assumed for the 6 MGD plant size. For the 30 and 100 MGD size plants, it was assumed that the combined primary and waste activated sludges would be disposed of by anaerobic digestion, vacuum filtration and incineration.

The excess activated sludge from both systems has been assumed to have a VSS/TSS ratio of 0.65. The reduction of VSS in the digestors has been assumed at 50 percent for the primary sludge, 50 percent for the air system waste activated sludge, and 40 percent for the oxygen system waste activated sludge. A lower percentage reduction of VSS for the oxygenation system waste activated sludge was assumed because of the somewhat higher level of oxidation that it has already undergone in the aeration tanks.

The quantity of excess activated sludge produced from each system was calculated using the correlation obtained from the data taken during the course of this study. These data have already been discussed in detail. The relationships used in each case are as follows:

For the Air System -
$$\frac{\text{Lbs excess VSS}}{\text{day}} = 1.38\ (\text{lbs. BOD removed/day}) - 0.174\ (\text{lbs. MLVSS under aeration})$$

For the Oxygen System -
$$\frac{\text{Lbs excess VSS}}{\text{day}} = 1.05\ (\text{lbs. BOD removed/day}) - 0.27\ (\text{lbs. MLVSS under aeration})$$

Table 43 summarizes the sludge disposal requirements for the air and oxygen systems for the 100 MGD plant size case for the various process design parameters. As shown, the amount of waste activated sludge generated in the oxygen system varies from about 28 to 57 percent of that produced in the air system for comparable levels of BOD removal. When this is added to the primary sludge produced in each case this amounts to about a 27 to 42 percent overall reduction in sludge disposal requirements for the oxygenation system.

TABLE 43

SUMMARY OF SLUDGE DISPOSAL REQUIREMENTS FOR AIR AND OXYGENATION SYSTEM ECONOMIC COMPARISONS

100 MGD PLANT

Aeration Detention Time, hrs. (Based on Raw Flow)	MLVSS Conc. (mg/1)	Lbs. MLVSS Under Aeration	Lbs. BOD Removed/Day	Primary Sludge Produced (lbs./Day)	Waste Activated Sludge Produced (lbs./Day)	Total Sludge Produced (Lbs./Day)
			AIR SYSTEM			
3.0	1625	169,000	150,000	86,000	148,000	234,000
4.0	1625	226,000	153,000	86,000	140,000	226,000
5.0	1625	282,000	156,000	86,000	131,000	217,000
6.0	1625	339,000	158,000	86,000	120,000	206,000
			OXYGENATION SYSTEM			
1.00	4420	154,000	150,000	86,000	84,000	170,000
1.33	4420	204,000	153,000	86,000	68,000	154,000
1.66	4420	255,000	156,000	86,000	51,000	137,000
2.00	4420	308,000	158,000	86,000	33,000	119,000

Capital Construction Costs

Investment costs were estimated for the aeration tanks, return sludge pumps and piping, secondary clarifiers, blowers and diffusers (including blower building, piping, and prime movers), the oxygenation dissolution equipment, and the oxygen supply generation equipment. Sludge disposal capital costs were not directly evaluated as will be discussed below. All investment costs except for oxygen supply and dissolution equipment, and covering of the aeration tanks were determined from information supplied directly by the FWQA. These costs reflect average January, 1970 construction costs in all cases. For specific locations and subsurface conditions, these costs could vary as much as plus or minus 40%.

Investment and total annual costs for other plant facilities common to both oxygen and air systems and unaffected by aeration techniques were supplied directly by the FWQA and are shown in Table 44. Included in these costs are raw sewage pumping, metering, preliminary treatment, primary sedimentation, service and administration buildings, and outside piping and appurtenances.

Figure 18 shows the bare aeration tank construction costs used as a function of the total aeration tank volume. These costs were used for both air and oxygenation systems. The additional costs of the aeration tank covers and staging baffles required for the oxygenation system were evaluated on the basis of an additional cost of $2.80 per square foot (6-inch thickness) of tank cover and staging baffle area required. Concrete construction is assumed for the tank covers and baffles.

Figure 19 shows the final sedimentation tank costs as a function of the surface area and tank depth. The final sedimentation basins were sized on the basis of an overflow rate of 800 GPD/ft^2 for both the air and oxygen systems. The basin depth used for the oxygen system was 11 feet as opposed to 10 feet for the air system. The slightly greater depth for the oxygen system was used to account for the greater concentration

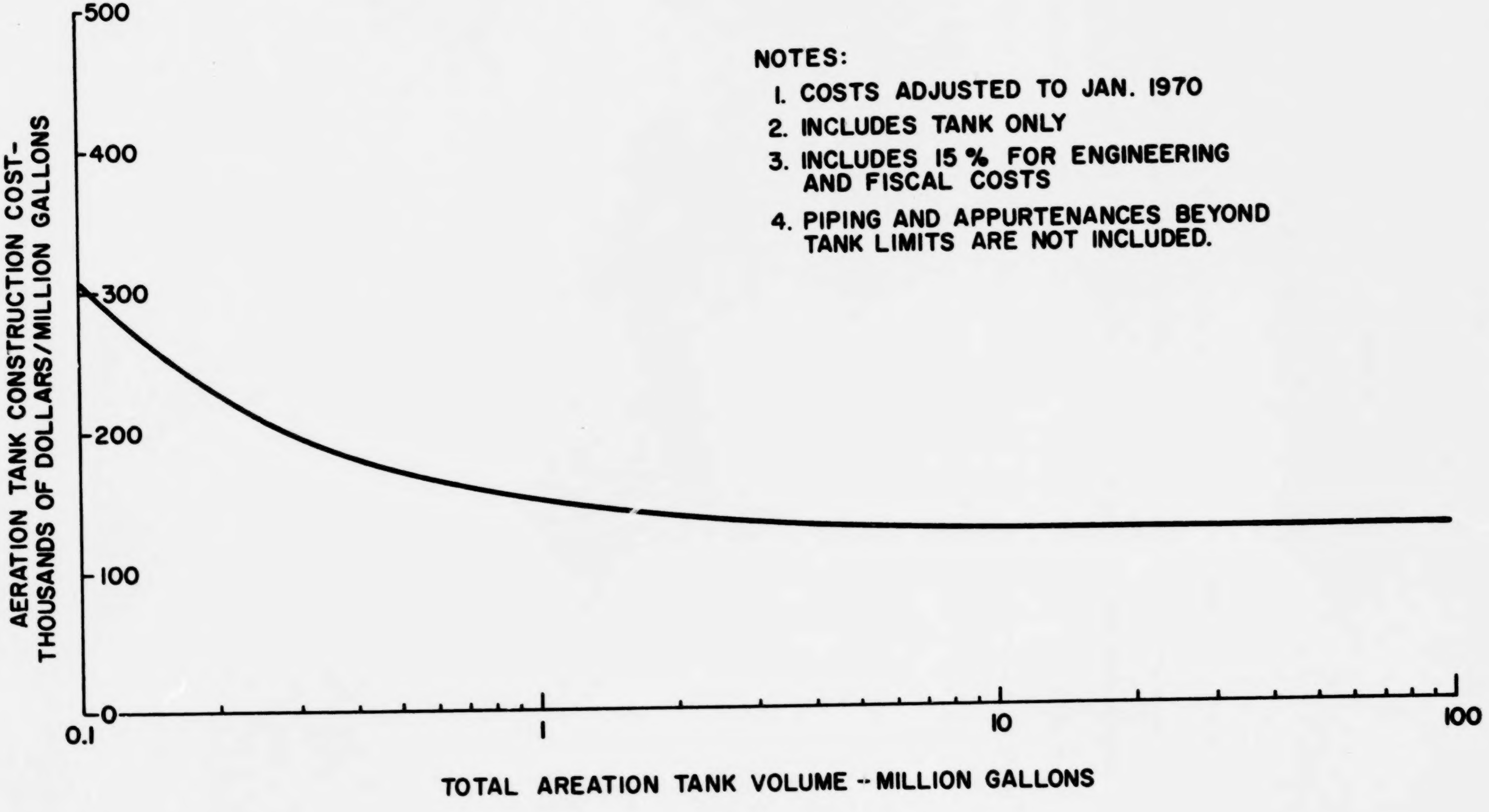
TABLE 44

COSTS FOR COMMON PLANT FACILITIES

Plant Size - MGD	1	6	30	100
Investment - $	190,000	600,000	2,000,000	5,400,000
Annual Investment Costs-$/Yr.	14,200	44,700	149,000	403,000
Annual O & M Costs - $/Yr.	12,300	34,200	110,000	302,000
Total Annual Costs - $/Yr.	26,500	78,900	259,000	705,000
Total Annual Costs-¢/1000 Gal.	7.3	3.6	2.4	1.9

FIGURE 18

AERATION TANK CONSTRUCTION COSTS

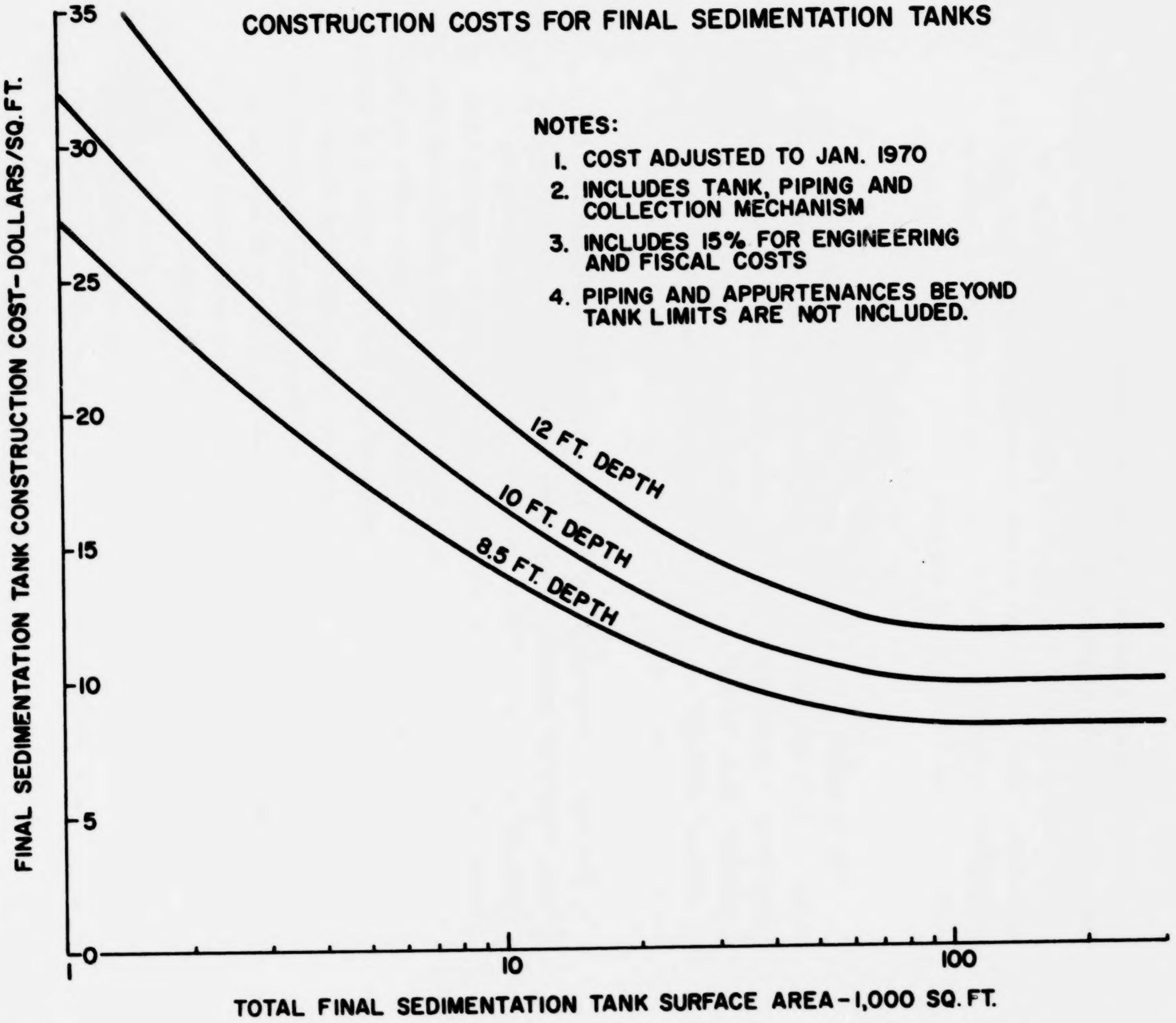


NOTES:

1. COSTS ADJUSTED TO JAN. 1970
2. INCLUDES TANK ONLY
3. INCLUDES 15 % FOR ENGINEERING AND FISCAL COSTS
4. PIPING AND APPURTENANCES BEYOND TANK LIMITS ARE NOT INCLUDED.

FIGURE 19

CONSTRUCTION COSTS FOR FINAL SEDIMENTATION TANKS



NOTES:

1. COST ADJUSTED TO JAN. 1970
2. INCLUDES TANK, PIPING AND COLLECTION MECHANISM
3. INCLUDES 15% FOR ENGINEERING AND FISCAL COSTS
4. PIPING AND APPURTENANCES BEYOND TANK LIMITS ARE NOT INCLUDED.

and hence inventory of solids that would exist in the oxygen system clarifier.

The air supply construction costs are shown in Figure 20 as a function of the air supply capacity in thousands of cfm. These costs include the blower building, blowers, prime movers, air piping and diffusers. These costs also include reserve or backup blower capacity as outlined in Table 45. Firm capacity is defined as the capacity with the largest unit out of service.

The secondary sludge recycle pumps and piping costs are shown in Figure 21 as a function of the sludge recycle pumping capacity in MGD. These costs include pump building, pumps, piping, motors, and associated electrical work. Sludge recycle rates are assumed to be 20 and 33 percent of influent flow for the oxygenation and air systems, respectively. Pumps are sized for a ratio of maximum to average plant flow of 2.0.

The capital cost estimates for the oxygenation equipment and the on-site oxygen gas generators are shown directly in the tabulated cost comparisons to follow. These figures represent installed costs for installation of the oxygenation equipment on the covered aeration tanks and the installation of the oxygen gas generators on a clear, level site in reasonably close proximity to the aeration tanks. These costs include all stainless steel piping, instrumentation and controls, electrical switchgear, motors, and etc. required for complete installation and automatic operation of the system. Capital costs for the oxygenation equipment and on-site oxygen gas generators include a one-time paid-up license fee to the Linde Division of Union Carbide Corporation for proprietary know-how and/or patent rights for which applications have been made.

The reserve or backup capacity supplied in the oxygen generators includes a complete spare air compression unit for the 1, 6, and 30 MGD plant sizes and a 24-hour storage of on-site liquid oxygen for the 100 MGD plants. In comparison to the reserve capacity supplied in the air blower

TABLE 45

RESERVE AIR BLOWER CAPACITY INCLUDED IN CONSTRUCTION COSTS OF AIR SUPPLY SYSTEM

Capacity, cfm	% Reserve Capacity	Total No. of Blowers
To 1,000	100	2
1,000 - 3,000	50-100	2 or 3
3,000 - 30,000	50	3
30,000 - 50,000	33 - 50	3 or 4
50,000 - 100,000	33	4
100,000 - 200,000	25 - 33	4 or 5
Over 200,000	20 - 25	5 or 6

FIGURE 20

AIR SUPPLY CONSTRUCTION COSTS

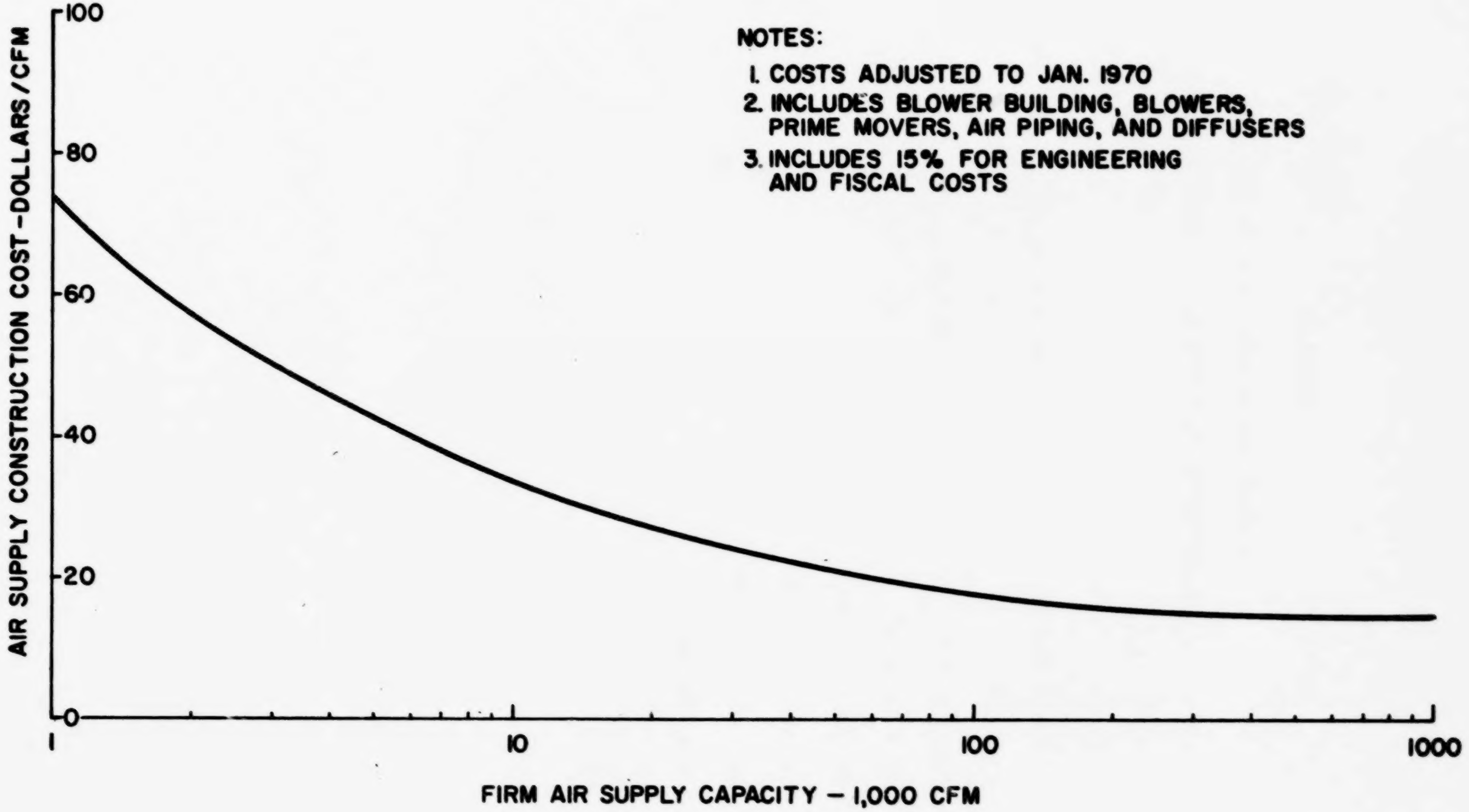
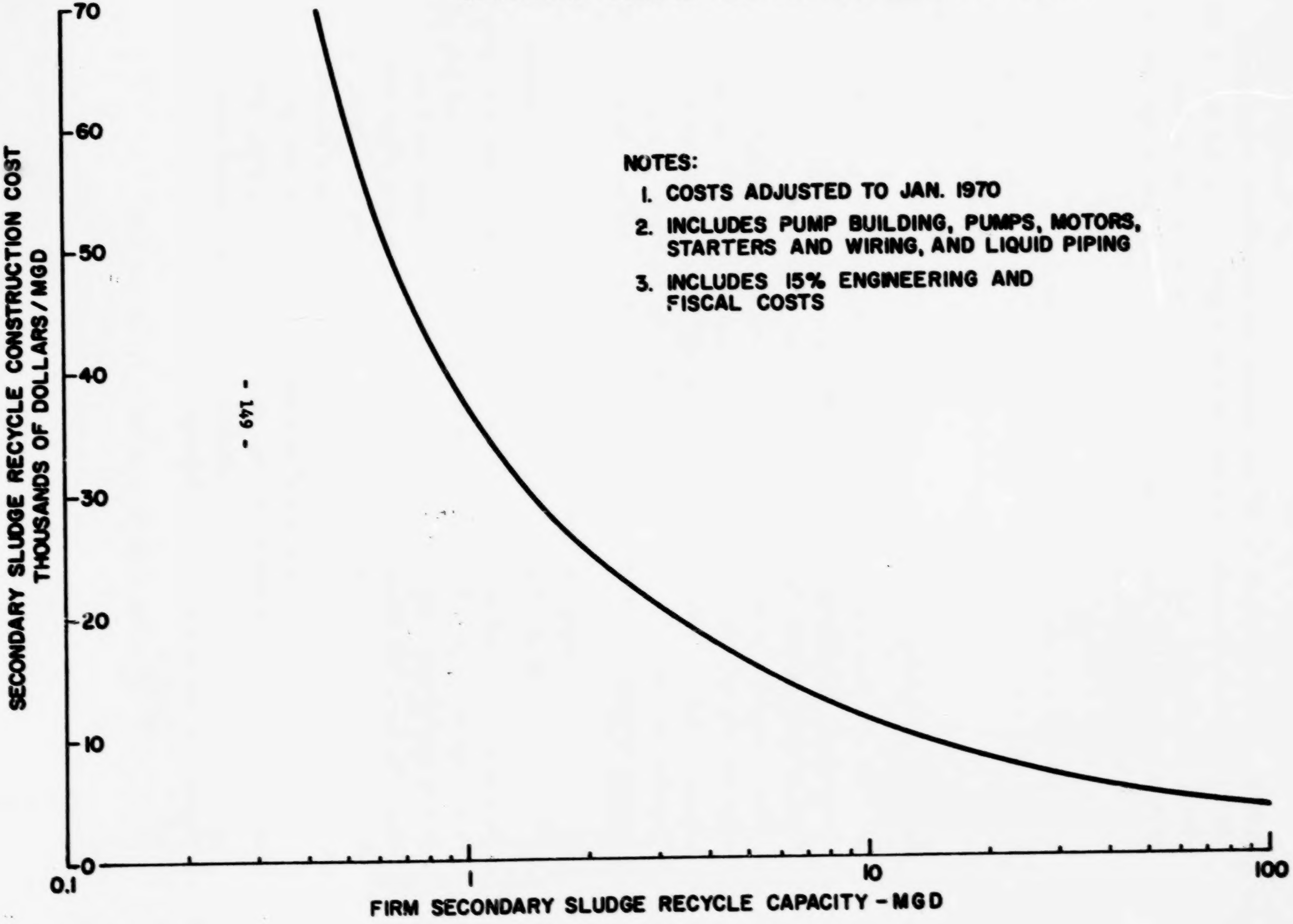


FIGURE 21

SECONDARY SLUDGE RECYCLE CONSTRUCTION COSTS



NOTES:

1. COSTS ADJUSTED TO JAN. 1970
2. INCLUDES PUMP BUILDING, PUMPS, MOTORS, STARTERS AND WIRING, AND LIQUID PIPING
3. INCLUDES 15% ENGINEERING AND FISCAL COSTS

systems, this corresponds to 100 per reserve air compression capacity for the 1, 6, and 30 MGD plant sizes and complete liquid back-up for the 100 MGD plant. The 24-hour liquid storage is sized on the basis that in most major municipalities, liquid oxygen could easily be supplied at the rate of 50 T/D on 24 hours notice.

All capital construction costs were amortized at a 5.5 percent interest rate over a period of 25 years. This corresponds to a capital recovery factor of 0.07455 or an annual amortization rate of 7.455 percent.

Operating and Maintenance Costs

The operating and maintenance costs for the secondary treatment facilities were obtained from a June, 1968, FWQA report entitled "Unit Process Operating and Maintenance Costs for Conventional Waste Treatment Plants", prepared by Mr. C. L. Swanson. The operating and maintenance costs for the aeration tanks, blowers, and final sedimentation basins of a concentional air aeration system were calculated from Mr. Swanson's correlations using the following relationship for 1970 costs for an aeration rate of 1.6 cf(NTP) of air per gallon of wastewater treated.

$$\$/\text{yr}/\text{MG} = 12{,}200 + 17{,}700\ (1/\text{MG})^{0.5}$$

where MG = volume of air aeration tanks in millions of gallons

The operating and maintenance costs for the air rates of 0.8 and 2.4 cf(NTP) of air per gallon were taken to be 0.9 and 1.1 times those above for 1.6 cf(NTP) per gallon. Of these costs it was assumed that about half was associated with the blowers and the final clarifiers and the other half was associated with the aeration tanks and diffusers. On this basis, it was decided that the operating and maintenance costs for the oxygenation system dissolution equipment would be comparable to that for the air blower system. Consequently, the operating and maintenance costs for the oxygenation equipment and final clarifiers of the oxygen system were calculated from the following relationship:

$$\$/\text{yr}/\text{EMG} = (0.5)\ \left(12{,}200 + 17{,}700\ (1/\ \text{EMG})^{0.5}\right)$$

where EMG = equivalent air aeration tank volume of oxygenation system in millions of gallons.

This procedure equates the operating and maintenance costs for the oxygenation equipment and final clarifier of the oxygen system to the blower equipment and final clarifier of the air system for a given size plant regardless of the relative system horsepower requirements or aeration tank sizes.

The operating and maintenance costs for the oxygenation system tanks were reduced by 25 percent on a per gallon basis because of the absence of diffusers, pipe headers, and etc. and the associated maintenance thereon. The operating and maintenance costs for the oxygenation tanks were then given by the following relationship:

$$\$/\text{yr}/\text{MG} = (0.375)\ (12{,}200 + 17{,}700\ (1/\text{MG})^{0.5})$$

where MG = volume of oxygenation tanks in millions of gallons.

The additional maintenance and repair costs associated with the oxygen generation plants were added to the above costs for the oxygen system. These costs are shown in Figure 22 as a function of the oxygen plant capacity in tons per day. These costs are representative of those incurred in Union Carbide owned oxygen generation facilities of comparable size and complexity.

Total net costs for oxygen gas generation including investment amortization (5.5 percent interest at 25 years), power, and maintenance and repair costs are shown in Figure 23 as a function of the plant generation capacity. Typical current costs for purchased oxygen gas from on-site gas generation facilities and purchased liquid oxygen are also shown for comparison purposes. As shown, the net cost is considerably lower for the purchased oxygen generation plant over the entire range of plant sizes. The primary reason for this is the lower cost of capital to a municipal government (5.5 percent interest rate over a 25 year period) and the inherently higher production and shipping cost of liquid oxygen.

The on-site oxygen gas generation facility for the 1, 6, and 30 MGD plants is of a design which has not heretofore been marketed on a broad scale for oxygen generation. These plants are based on sound, well proven engineering design principles, however, which have been thoroughly tested and employed in a variety of gas separation processes for a number of years. These units have operated automatically and unattended for several years without requiring any maintenance or repair. The oxygen plant maintenance and repair costs shown in Figure 22 represent an annual cost of about two percent of the original capital costs and represent a conservative estimate of these costs based on many years of operating experience.

Electric power costs for air aeration, oxygen dissolution, oxygen generation, and recycle sludge pumping were evaluated on the basis of the following rate schedule:

Plant Size (MGD)	Electric Utility Rate ($/KWH)
1	0.0150
6	0.0118
30	0.00943
100	0.00800

The recycle sludge pump head capacity was taken as the depth of the aeration tanks plus 14 feet. The sludge pump efficiency was assumed at 80 percent.

The compressor efficiencies assumed for the air blowers and the oxygenation system gas recirculation compressors were 70 and 74% and all motor efficiencies were assumed at 95 percent. These various unit equipment efficiencies were then used with the above power rate schedules to arrive at the electric utility costs for the two processes at the various plant sizes.

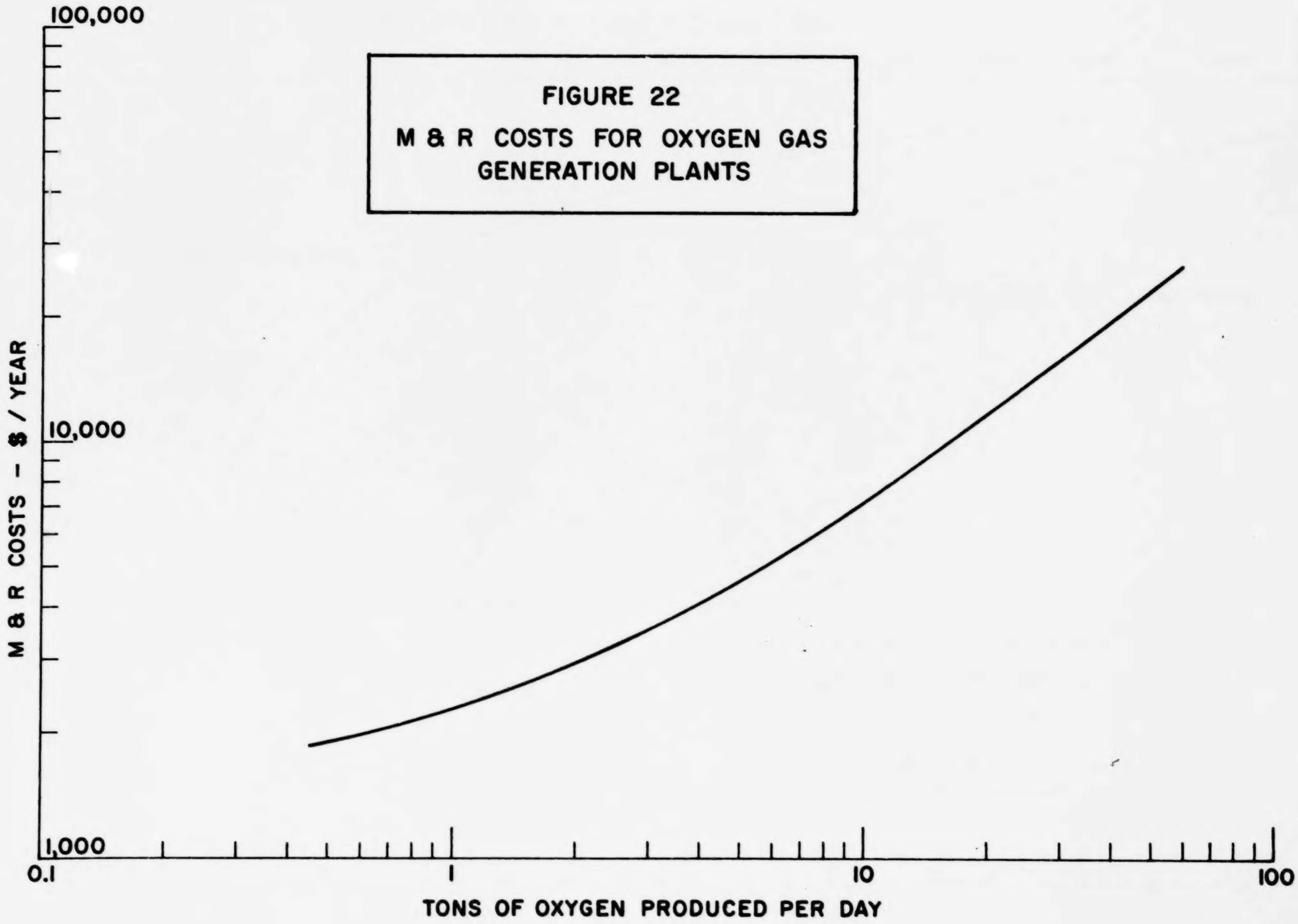


FIGURE 22

M & R COSTS FOR OXYGEN GAS GENERATION PLANTS

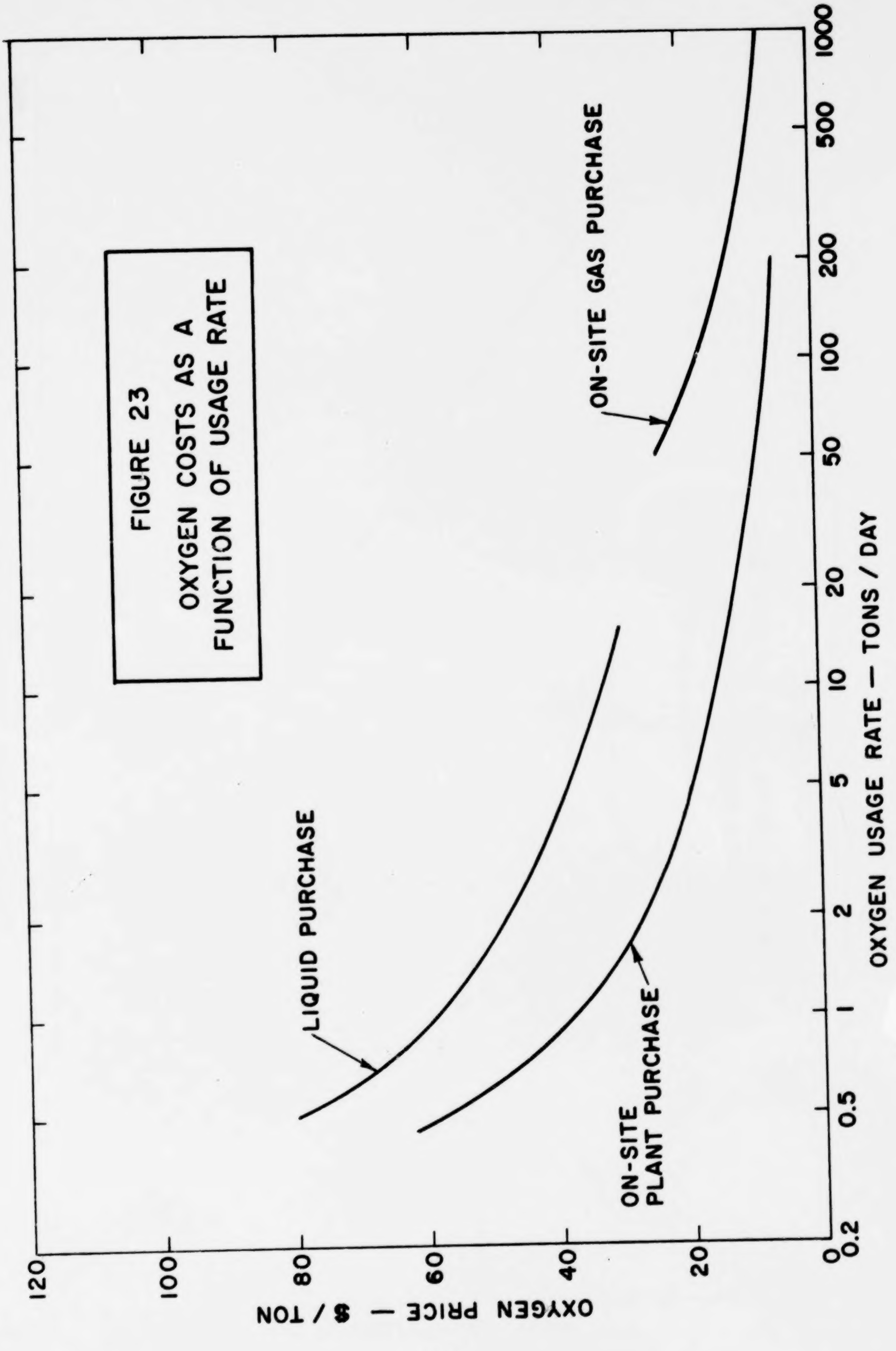


FIGURE 23

OXYGEN COSTS AS A FUNCTION OF USAGE RATE

Sludge Disposal Costs

As stated previously, sludge disposal for the 1 MGD plant size was assumed to be by anaerobic digestion and sludge drying beds while that for the 6 MGD plant was assumed to be by anaerobic digestion, vacuum filtration, and cake disposal by land fill. Sludge disposal for the 30 and 100 MGD plant sizes was assumed to be anaerobic digestion, vacuum filtration, and incineration. In addition, gravity thickening is required for the air system waste activated sludge regardless of the assumed subsequent disposal methods. Primary sludge and oxygen system waste activated sludge are not assumed to require thickening. Costs associated with these disposal methods also were supplied by the FWQA as total system costs including amortization and operating and maintenance costs. Amortization costs were calculated at 5.5 percent for 25 years and all capital costs were adjusted to January, 1970, conditions.

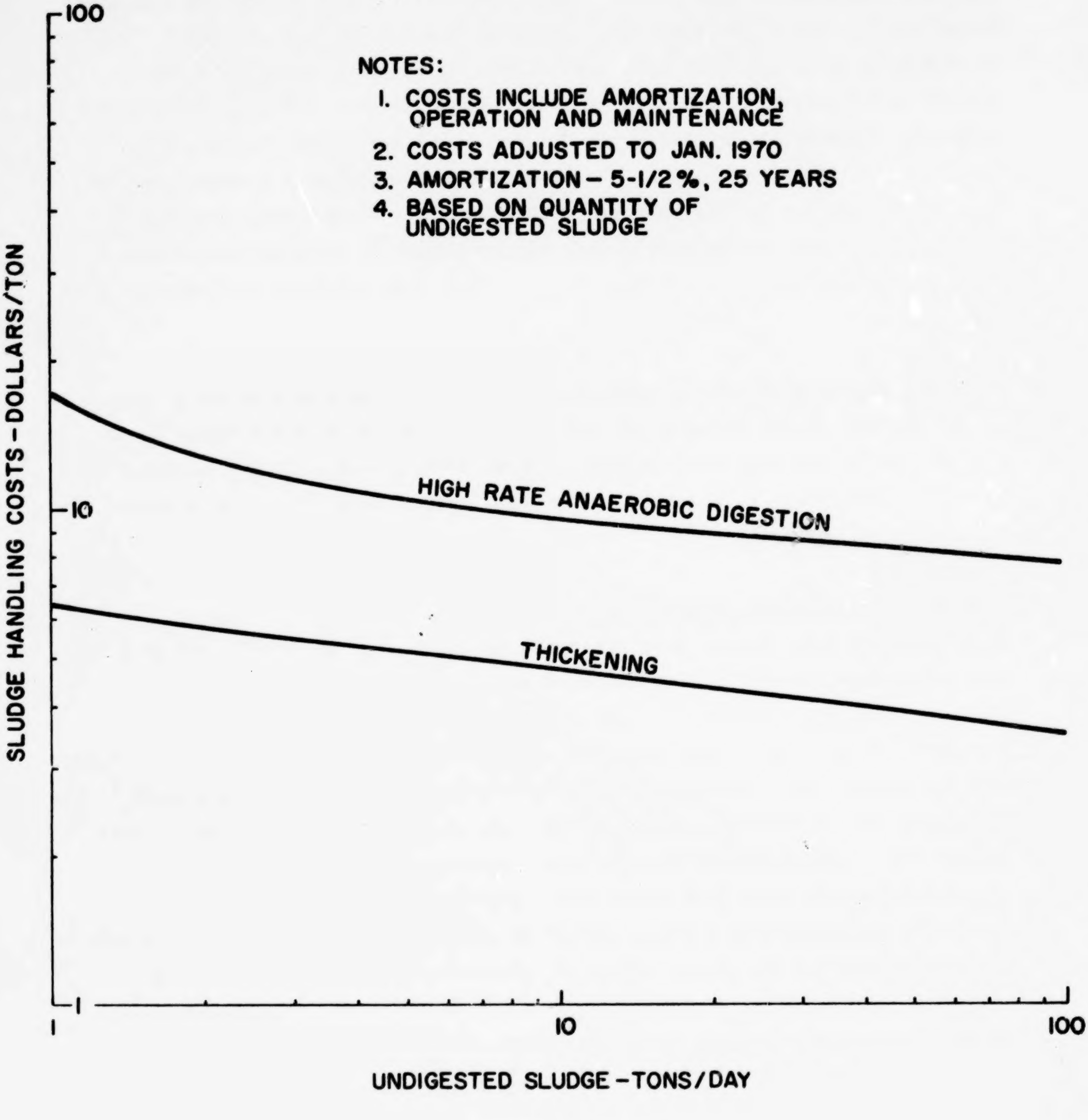
The anaerobic digestion and sludge thickening costs are shown in Figure 24 as a function of the quantity of undigested sludge to be processed. Costs for vacuum filtration, incineration, sludge drying beds and cake disposal by land fill are shown in Figure 25 as a function of the quantity of digested sludge to be processed.

Discussion of Economic Comparison

Detailed cost comparisons are contained in Tables 46 through 61 wherein the investment, operating and sludge disposal costs are evaluated for each of four plant sizes and all of the process conditions outlined previously for both the air and oxygen systems. Also included in these cost comparisons are the annual costs associated with facilities not related to secondary treatment and, therefore, common to both the air and oxygen systems. Total annual costs associated with secondary treatment in each case are arrived at by summing the amortized investment costs at 5.5% and 25 years and the operating and maintenance costs including power. The addition of the sludge disposal costs and the annual costs of common facilities provides total treatment costs. Total treatment system annual costs are then reduced to cents per thousand gallons treated in each case.

FIGURE 24

COSTS OF SLUDGE HANDLING PROCESSES



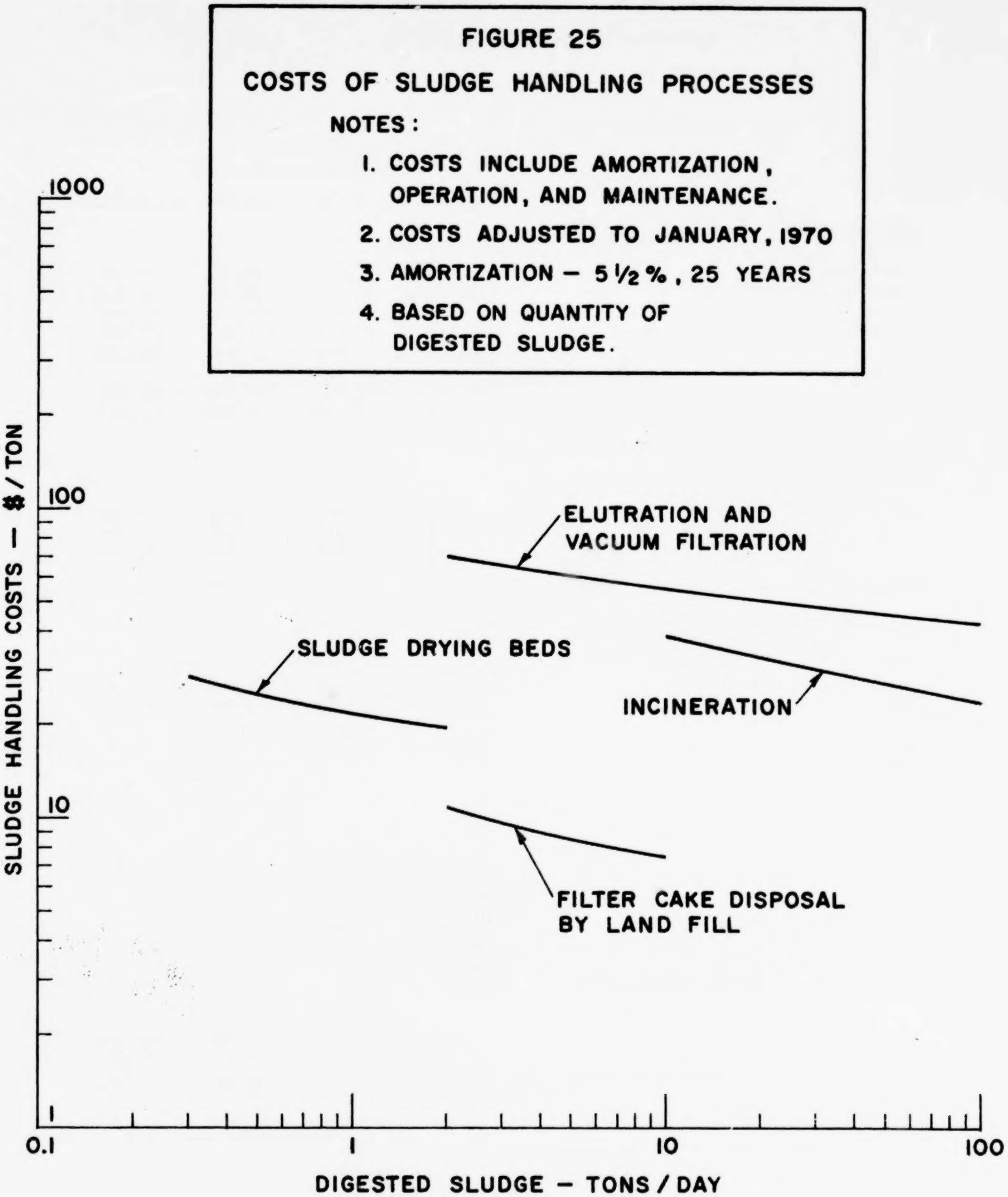


FIGURE 25

COSTS OF SLUDGE HANDLING PROCESSES

NOTES:

1. COSTS INCLUDE AMORTIZATION, OPERATION, AND MAINTENANCE.
2. COSTS ADJUSTED TO JANUARY, 1970
3. AMORTIZATION — 5 1/2 %, 25 YEARS
4. BASED ON QUANTITY OF DIGESTED SLUDGE.

TABLE 46

ESTIMATED SECONDARY TREATMENT, SLUDGE DISPOSAL, AND TOTAL PLANT COSTS WITH OXYGENATION

1 MGD Treated

	Hours Detention Based on Raw Sewage Flow			
	1.00	1.33	1.66	2.00
SECONDARY TREATMENT SYSTEM INVESTMENT, $				
Oxygenation Tanks	23,000	26,000	29,000	32,000
Sludge Return Pumps & Piping	30,000	30,000	30,000	30,000
Oxygen Dissolution Equipment	103,000	104,000	105,000	106,000
Oxygen Generation Equipment	99,000	101,000	103,000	105,000
Final Sedimentation Tanks	41,000	41,000	41,000	41,000
Total Secondary Investment	296,000	302,000	308,000	314,000
Annual Secondary Investment Cost, $/Yr.	22,100	22,500	23,000	23,400
SECONDARY TREATMENT SYSTEM OPERATING & MAINTENANCE COST, $/YEAR				
Oxygen Generation Power	1,100	1,300	1,400	1,500
Oxygen Dissolution Power	700	800	900	1,000
M & R Costs & Sludge Return Pump Power	8,300	9,300	10,200	11,200
Annual Secondary O & M Costs	10,100	11,400	12,500	13,700
SECONDARY TREATMENT SYSTEM ANNUAL COSTS, $/YEAR				
Secondary Investment	22,100	22,500	23,000	23,400
Secondary O & M	10,100	11,400	12,500	13,700
Annual Secondary Treatment Costs	32,200	33,900	35,500	37,100
Cents/1000 gallons	8.8	9.3	9.7	10.2
TOTAL PRIMARY & SECONDARY SLUDGE DISPOSAL COSTS, $/YEAR				
Anaerobic Digestion	5,900	5,600	5,400	5,200
Sludge Drying Bed Disposal	5,200	4,800	4,400	3,900
Annual Sludge Disposal Costs	11,100	10,400	9,800	9,100
Cents/1000 Gallons	3.0	2.8	2.7	2.5
TOTAL TREATMENT SYSTEM ANNUAL COSTS, $/YEAR				
Annual Secondary Costs	32,200	33,900	35,500	37,100
Annual Sludge Disposal Costs	11,100	10,400	9,800	9,100
Annual Common Plant Facilities Costs	26,500	26,500	26,500	26,500
Total Plant Annual Costs	69,800	70.800	71,800	72,700
Cents/1000 Gallons	19.1	19.4	19.7	19.9

TABLE 47

ESTIMATED SECONDARY TREATMENT, SLUDGE DISPOSAL, AND TOTAL PLANT COSTS WITH OXYGENATION

6 MGD Treated

	Hours Detention Based on Raw Sewage Flow			
	1.00	1.33	1.66	2.00
SECONDARY TREATMENT SYSTEM INVESTMENT,$				
Oxygenation Tanks	62,000	75,000	89,000	103,000
Sludge Return Pumps & Piping	56,000	56,000	56,000	56,000
Oxygen Dissolution Equipment	155,000	159,000	167,000	175,000
Oxygen Generation Equipment	200,000	210,000	218,000	227,000
Final Sedimentation Tanks	147,000	147,000	147,000	147,000
Total Secondary Investment	620,000	647,000	677,000	708,000
Annual Secondary Investment Costs,$/Yr.	46,200	48,200	50,500	52,800
SECONDARY TREATMENT SYSTEM OPERATING & MAINTENANCE COSTS, $/YEAR				
Oxygen Generation Power	5,100	5,700	6,300	6,900
Oxygen Dissolution Power	3,000	3,500	4,000	4,600
M & R Costs & Sludge Return Pump Power	21,600	25,300	28,900	33,300
Annual Secondary O & M Costs	29,700	34,500	39,200	44,800
SECONDARY TREATMENT SYSTEM ANNUAL COSTS, $/YEAR				
Secondary Investment	46,200	48,200	50,500	52,800
Secondary O & M	29,700	34,500	39,200	44,800
Annual Secondary Treatment Costs	75,900	82,700	89,700	97,600
Cents/1000 Gallons	3.5	3.8	4.1	4.5
TOTAL PRIMARY & SECONDARY SLUDGE DISPOSAL COSTS, $/YEAR				
Anaerobic Digestion	19,100	17,700	16,200	15,800
Vacuum Filtration	82,500	76,700	67,700	60,100
Cake Disposal by Landfill	11,900	11,100	10,200	9,000
Annual Sludge Disposal Costs	113,500	105,500	94,100	84,900
Cents/1000 Gallons	5.2	4.8	4.3	3.9
TOTAL TREATMENT SYSTEM ANNUAL COSTS, $/YEAR				
Annual Secondary Costs	75,900	82,700	89,700	97,600
Annual Sludge Disposal Costs	113,500	105,500	94,100	84,900
Annual Common Plant Facilities Costs	78,900	78,900	78,900	78,900
Total Plant Annual Costs	268,300	267,100	262,700	261,400
Cents/1000 Gallons	12.3	12.2	12.0	11.9

TABLE 48

ESTIMATED SECONDARY TREATMENT, SLUDGE DISPOSAL, AND TOTAL PLANT COSTS WITH OXYGENATION

30 MGD Treated

	Hours Detention Based on Raw Sewage Flow			
	1.00	1.33	1.66	2.00
SECONDARY TREATMENT SYSTEM INVESTMENT,$				
Oxygenation Tanks	215,000	274,000	337,000	394,000
Sludge Return Pumps & Piping	121,000	121,000	121,000	121,000
Oxygen Dissolution Equipment	471,000	503,000	547,000	571,000
Oxygen Generation Equipment	595,000	635,000	670,000	705,000
Final Sedimentation Tanks	467,000	467,000	467,000	467,000
Total Secondary Investment	1,869,000	2,000,000	2,142,000	2,258,000
Annual Secondary Investment Cost,$/Yr.	139,000	149,000	160,000	168,000
SECONDARY TREATMENT SYSTEM OPERATING & MAINTENANCE COSTS, $/YEAR				
Oxygen Generation Power	19,000	22,000	25,000	27,000
Oxygen Dissolution Power	11,000	14,000	16,000	18,000
M & R Costs & Sludge Return Pump Power	63,000	77,000	90,000	104,000
Annual Secondary O & M Costs	93,000	113,000	131,000	149,000
SECONDARY TREATMENT SYSTEM ANNUAL COSTS, $/YEAR				
Secondary Investment	139,000	149,000	160,000	168,000
Secondary O & M	93,000	113,000	131,000	149,000
Annual Secondary Treatment Costs	232,000	262,000	291,000	317,000
Cents/1000 Gallons	2.1	2.4	2.7	2.9
TOTAL PRIMARY & SECONDARY SLUDGE DISPOSAL COSTS, $/YEAR				
Anaerobic Digestion	82,000	75,000	68,000	59,000
Vacuum Filtration	334,000	305,000	273,000	238,000
Incineration	222,000	203,000	185,000	166,000
Annual Sludge Disposal Costs	638,000	583,000	526,000	463,000
Cents/1000 Gallons	5.8	5.3	4.8	4.2
TOTAL TREATMENT SYSTEM ANNUAL COSTS, $/YEAR				
Annual Secondary Costs	232,000	262,000	291,000	317,000
Annual Sludge Disposal Costs	638,000	583,000	526,000	463,000
Annual Common Plant Facilities Costs	259,000	259,000	259,000	259,000
Total Plant Annual Costs	1,129,000	1,104,000	1,076,000	1,039,000
Cents/1000 Gallons	10.3	10.1	9.8	9.5

TABLE 49

ESTIMATED SECONDARY TREATMENT, SLUDGE DISPOSAL, AND TOTAL PLANT COSTS WITH OXYGENATION

100 MGD Treated

	Hours Detention Based on Raw Sewage Flow			
	1.00	1.33	1.66	2.00
SECONDARY TREATMENT SYSTEM INVESTMENT,$				
Oxygenation Tanks	631,000	817,000	1,017,000	1,217,000
Sludge Return Pumps & Piping	236,000	236,000	236,000	236,000
Oxygen Dissolution Equipment	1,678,000	1,753,000	1,835,000	1,978,000
Oxygen Generation Equipment	1,090,000	1,150,000	1,210,000	1,280,000
Final Sedimentation Tanks	1,350,000	1,350,000	1,350,000	1,350,000
Total Secondary Investment	4,985,000	5,306,000	5,648,000	6,061,000
Annual Secondary Investment Costs,$/Yr.	372,000	396,000	421,000	452,000
SECONDARY TREATMENT SYSTEM OPERATING & MAINTENANCE COSTS, $/YEAR				
Oxygen Generation Power	44,000	50,000	55,000	60,000
Oxygen Dissolution Power	28,000	34,000	41,000	48,000
M & R Costs & Sludge Return Pump Power	170,000	211,000	248,000	286,000
Annual Secondary O & M Costs	242,000	295,000	344,000	394,000
SECONDARY TREATMENT SYSTEM ANNUAL COSTS, $/YEAR				
Secondary Investment	372,000	396,000	421,000	452,000
Secondary O & M	242,000	295,000	344,000	394,000
Annual Secondary Treatment Costs	614,000	691,000	765,000	846,000
Cents/1000 Gallons	1.7	1.9	2.1	2.3
TOTAL PRIMARY & SECONDARY SLUDGE DISPOSAL COSTS, $/YEAR				
Anaerobic Digestion	244,000	225,000	202,000	179,000
Vacuum Filtration	966,000	893,000	796,000	691,000
Incineration	569,000	523,000	476,000	419,000
Annual Sludge Disposal Costs	1,779,000	1,640,000	1,474,000	1,289,000
Cents/1000 Gallons	4.9	4.5	4.0	3.5
TOTAL TREATMENT SYSTEM ANNUAL COSTS, $/YEAR				
Annual Secondary Costs	614,000	691,000	765,000	846,000
Annual Sludge Disposal Costs	1,779,000	1,640,000	1,474,000	1,289,000
Annual Common Plant Facilities Costs	705,000	705,000	705,000	705,000
Total Plant Annual Costs	3,098,000	3,036,000	2,944,000	2,840,000
Cents/1000 Gallons	8.5	8.3	8.1	7.8

TABLE 50

ESTIMATED SECONDARY TREATMENT, SLUDGE DISPOSAL, AND TOTAL PLANT COSTS WITH AIR AERATION

1 MGD Treated, 0.8 CF(NTP) of Air/Gallon, or 750 CF(NTP) of Air/Lb. BOD

	Hours Detention Based on Raw Sewage Flow			
	3.0	4.0	5.0	6.0
SECONDARY TREATMENT SYSTEM INVESTMENT, $				
Aeration Tanks	35,000	41,000	46,000	52,000
Sludge Return Pumps & Piping	33,000	33,000	33,000	33,000
Blowers & Diffusers	59,000	59,000	59,000	59,000
Final Sedimentation Tanks	38,000	38,000	38,000	38,000
Total Secondary Investment	165,000	171,000	176,000	182,000
Annual Secondary Investment Cost,$/yr.	12,300	12,700	13,100	13,600
SECONDARY TREATMENT SYSTEM OPERATING & MAINTENANCE COSTS, $/YEAR				
Blower Power	2,500	2,500	2,500	2,500
M & R Costs & Sludge Return Pump Power	7,200	8,600	9,800	10,900
Annual Secondary O & M Costs	9,700	11,100	12,300	13,400
SECONDARY TREATMENT SYSTEM ANNUAL COSTS, $/YEAR				
Secondary Investment	12,300	12,700	13,100	13,600
Secondary O & M	9,700	11,100	12,300	13,400
Annual Secondary Treatment Costs	22,000	23,800	25,400	27,000
Cents/1000 gallons	6.0	6.5	7.0	7.4
TOTAL PRIMARY & SECONDARY SLUDGE DISPOSAL COSTS $/YEAR				
Waste Activated Sludge Thickening	1,900	1,800	1,800	1,700
Anaerobic Digestion	6,700	6,600	6,400	6,300
Sludge Drying Bed Disposal	6,500	6,300	6,100	5,800
Annual Sludge Disposal Costs	15,100	14,700	14,300	13,800
Cents/1000 Gallons	4.1	4.0	3.9	3.8
TOTAL TREATMENT SYSTEM ANNUAL COSTS, $/YEAR				
Annual Secondary Costs	22,000	23,800	25,400	27,000
Annual Sludge Disposal Costs	15,100	14,700	14,300	13,800
Annual Common Plant Facilities Costs	26,500	26,500	26,500	26,500
Total Plant Annual Costs	63,600	65,000	66,200	67,300
Cents/gallons	17.4	17.8	18.1	18.4

TABLE 51

ESTIMATED SECONDARY TREATMENT, SLUDGE DISPOSAL, AND TOTAL PLANT COSTS WITH AIR AERATION

1 MGD Treated, 1.6 CF(NTP) of Air/Gallon or 1500 CF(NTP) of Air/Lb. BOD

	Hours Detention Based on Raw Sewage Flow			
	3.0	4.0	5.0	6.0
SECONDARY TREATMENT SYSTEM INVESTMENT,$				
Aeration Tanks	35,000	41,000	46,000	52,000
Sludge Return Pumps & Piping	33,000	33,000	33,000	33,000
Blowers & Diffusers	79,000	79,000	79,000	79,000
Final Sedimentation Tanks	38,000	38,000	38,000	38,000
Total Secondary Investment	185,000	191,000	196,000	202,000
Annual Secondary Investment Costs,$/Yr.	13,800	14,200	14,600	15,100
SECONDARY TREATMENT SYSTEM OPERATING & MAINTENANCE COSTS, $/YEAR				
Blower Power	5,100	5,100	5,100	5,100
M & R Costs & Sludge Return Pump Power	8,000	9,500	10,800	12,100
Annual Secondary O & M Costs	13,100	14,600	15,900	17,200
SECONDARY TREATMENT SYSTEM ANNUAL COSTS, $/YEAR				
Secondary Investment	13,800	14,200	14,600	15,100
Secondary O & M	13,100	14,600	15,900	17,200
Annual Secondary Treatment Costs	26,900	28,800	30,500	32,300
Cents/1000 gallons	7.4	7.9	8.4	8.8
TOTAL PRIMARY & SECONDARY SLUDGE DISPOSAL COSTS, $/YEAR				
Waste Activated Sludge Thickening	1,900	1,800	1,800	1,700
Anaerobic Digestion	6,700	6,600	6,400	6,300
Sludge Drying Bed Disposal	6,500	6,300	6,100	5,800
Annual Sludge Disposal Costs	15,100	14,700	14,300	13,800
Cents/1000 gallons	4.1	4.0	3.9	3.8
TOTAL TREATMENT SYSTEM ANNUAL COSTS, $/YEAR				
Annual Secondary Costs	26,900	28,800	30,500	32,300
Annual Sludge Disposal Costs	15,100	14,700	14,300	13,800
Annual Common Plant Facilities Costs	26,500	26,500	26,500	26,500
Total Plant Annual Costs	68,500	70,000	71,300	72,600
Cents/1000 gallons	18.8	19.2	19.5	19.9

TABLE 52

ESTIMATED SECONDARY TREATMENT, SLUDGE DISPOSAL, AND TOTAL PLANT COSTS WITH AIR AERATION

1 MGD Treated, 2.4 CF(NTP) of Air/Gallon, or 2250 CF(NTP) of Air/Lb. BOD

	Hours Detention Based on Raw Sewage Flow			
	3.0	4.0	5.0	6.0
SECONDARY TREATMENT SYSTEM INVESTMENT, $				
Aeration Tanks	35,000	41,000	46,000	52,000
Sludge Return Pumps & Piping	33,000	33,000	33,000	33,000
Blowers & Diffusers	102,000	102,000	102,000	102,000
Final Sedimentation Tanks	38,000	38,000	38,000	38,000
Total Secondary Investment	208,000	214,000	219,000	225,000
Annual Secondary Investment Costs,$/Yr.	15,500	16,000	16,300	16,800
SECONDARY TREATMENT SYSTEM OPERATING & MAINTENANCE COSTS, $/YEAR				
Blower Power	7,600	7,600	7,600	7,600
M & R Costs & Sludge Return Pump Power	8,800	10,400	11,900	13,300
Annual Secondary O & M Costs	16,400	18,000	19,500	20,900
SECONDARY TREATMENT SYSTEM ANNUAL COSTS $/YEAR				
Secondary Investment	15,500	16,000	16,300	16,800
Secondary O & M	16,400	18,000	19,500	20,900
Annual Secondary Treatment Costs	31,900	34,000	35,800	37,700
Cents/1000 gallons	8.7	9.3	9.8	10.3
TOTAL PRIMARY & SECONDARY SLUDGE DISPOSAL COSTS, $/YEAR				
Waste Activated Sludge Thickening	1,900	1,800	1,800	1,700
Anaerobic Digestion	6,700	6,600	6,400	6,300
Sludge Drying Bed Disposal	6,500	6,300	6,100	5,800
Annual Sludge Disposal Costs	15,100	14,700	14,300	13,800
Cents/1000 gallons	4.1	4.0	3.9	3.8
TOTAL TREATMENT SYSTEM ANNUAL COSTS, $/YEAR				
Annual Secondary Costs	31,900	34,000	35,800	37,700
Annual Sludge Disposal Costs	15,100	14,700	14,300	13,800
Annual Common Plant Facilities Costs	26,500	26,500	26,500	26,500
Total Plant Annual Costs	73,500	75,200	76,600	78,000
Cents/1000 gallons	20.1	20.6	21.0	21.4

TABLE 53

ESTIMATED SECONDARY TREATMENT, SLUDGE DISPOSAL, AND TOTAL PLANT COSTS WITH AIR AERATION

6 MGD Treated, 0.8 CF(NTP) of Air/Gallon, or 750 CF(NTP) of Air/Lb. BOD

	Hours Detention Based on Raw Sewage Flow			
	3.0	4.0	5.0	6.0
SECONDARY TREATMENT SYSTEM INVESTMENT, $				
Aeration Tanks	118,000	151,000	181,000	213,000
Sludge Return Pumps & Piping	71,000	71,000	71,000	71,000
Blowers & Diffusers	161,000	161,000	161,000	161,000
Final Sedimentation Tanks	134,000	134,000	134,000	134,000
Total Secondary Investment	484,000	517,000	547,000	579,000
Annual Secondary Investment Costs,$/Yr.	36,100	38,500	40,800	43,200
SECONDARY TREATMENT SYSTEM OPERATING & MAINTENANCE COSTS $/YEAR				
Blower Power	11,900	11,900	11,900	11,900
M & R Costs & Sludge Return Pumps Power	23,100	28,100	32,600	35,900
Annual Secondary O & M Costs	35,000	40,000	44,500	47,800
SECONDARY TREATMENT SYSTEM ANNUAL COSTS, $/YEAR				
Secondary Investment	36,100	38,500	40,800	43,200
Secondary O & M	35,000	40,000	44,500	47,800
Annual Secondary Treatment Costs	71,100	78,500	85,300	91,000
Cents/1000 gallons	3.2	3.6	3.9	4.2
TOTAL PRIMARY & SECONDARY SLUDGE DISPOSAL COSTS, $/YEAR				
Waste Activated Sludge Thickening	8,600	8,100	7,700	7,100
Anaerobic Digestion	25,200	24,500	23,600	22,500
Vacuum Filtration	104,000	101,000	97,800	92,800
Cake Disposal by Landfill	14,700	14,400	14,000	13,500
Annual Sludge Disposal Costs	152,500	148,000	143,100	135,900
Cents/1000 gallons	7.0	6.8	6.5	6.2
TOTAL TREATMENT SYSTEM ANNUAL COSTS, $/YEAR				
Annual Secondary Costs	71,100	78,500	85,300	91,000
Annual Sludge Disposal Costs	152,500	148,000	143,100	135,900
Annual Common Plant Facilities Costs	78,900	78,900	78,900	78,900
Total Plant Annual Costs	302,500	305,400	307,300	305,800
Cents/1000 gallons	13.8	13.9	14.0	14.0

TABLE 54

ESTIMATED SECONDARY TREATMENT, SLUDGE DISPOSAL, AND TOTAL PLANT COSTS WITH AIR AERATION

6 MGD Treated, 1.6 CF(NTP) of Air/Gallon, or 1500 CF(NTP) of Air/Lb. BOD

	Hours Detention Based on Raw Sewage Flow			
	3.0	4.0	5.0	6.0
SECONDARY TREATMENT SYSTEM INVESTMENT, $				
Aeration Tanks	118,000	151,000	181,000	213,000
Sludge Return Pumps & Piping	71,000	71,000	71,000	71,000
Blowers & Diffusers	257,000	257,000	257,000	257,000
Final Sedimentation Tanks	134,000	134,000	134,000	134,000
Total Secondary Investment	580,000	613,000	643,000	675,000
Annual Secondary Investment Costs, $/Yr.	43,200	45,700	47,900	50,300
SECONDARY TREATMENT SYSTEM OPERATING & MAINTENANCE COSTS, $/YEAR				
Blower Power	23,800	23,800	23,800	23,800
M & R Costs & Sludge Return Pump Power	25,600	31,000	36,200	41,300
Annual Secondary O & M Costs	49,400	54,800	60,000	65,100
SECONDARY TREATMENT SYSTEM ANNUAL COSTS, $/YEAR				
Secondary Investment	43,200	45,700	47,900	50,300
Secondary O & M	49,400	54,800	60,000	65,100
Annual Secondary Treatment Costs	92,600	100,500	107,900	115,400
Cents/1000 gallons	4.2	4.6	4.9	5.3
TOTAL PRIMARY & SECONDARY SLUDGE DISPOSAL COSTS, $/YEAR				
Waste Activated Sludge Thickening	8,600	8,100	7,700	7,100
Anaerobic Digestion	25,200	24,500	23,600	22,500
Vacuum Filtration	104,000	101,000	97,800	92,800
Cake Disposal by Landfill	14,700	14,400	14,000	13,500
Annual Sludge Disposal Costs	152,500	148,000	143,100	135,900
Cents/1000 Gallons	7.0	6.8	6.5	6.2
TOTAL TREATMENT SYSTEM ANNUAL COSTS, $/YEAR				
Annual Secondary Costs	92,600	100,500	107,900	115,400
Annual Sludge Disposal Costs	152,500	148,000	143,100	135,900
Annual Common Plant Facilities Costs	78,900	78,900	78,900	78,900
Total Plant Annual Costs	324,000	327,400	329,900	330,200
Cents/1000 gallons	14.8	14.9	15.1	15.1

TABLE 55

ESTIMATED SECONDARY TREATMENT, SLUDGE DISPOSAL, AND TOTAL PLANT COSTS WITH AIR AERATION

6 MGD Treated, 2.4 CF(NTP) of Air/Gallon, or 2250 CF(NTP) of Air/Lb. BOD

	Hours Detention Based on Raw Sewage Flow			
	3.0	4.0	5.0	6.0
SECONDARY TREATMENT SYSTEM INVESTMENT, $				
Aeration Tanks	118,000	151,000	181,000	213,000
Sludge Return Pumps & Piping	71,000	71,000	71,000	71,000
Blowers & Diffusers	340,000	340,000	340,000	340,000
Final Sedimentation Tanks	134,000	134,000	134,000	134,000
Total Secondary Investment	663,000	696,000	726,000	758,000
Annual Secondary Investment Cost, $/Yr.	49,400	51,900	54,000	56,500
SECONDARY TREATMENT SYSTEM OPERATING & MAINTENANCE COSTS, $/YEAR				
Blower Power	35,700	35,700	35,700	35,700
M & R Costs & Sludge Return Pump Power	27,900	34,000	39,600	45,000
Annual Secondary O & M Costs	63,600	69,700	75,300	80,700
SECONDARY TREATMENT SYSTEM ANNUAL COSTS, $/YEAR				
Secondary Investment	49,400	51,900	54,100	56,500
Secondary O & M	63,600	69,700	75,300	80,700
Annual Secondary Treatment Costs	113,000	121,600	129,400	137,200
Cents/1000 gallons	5.1	5.6	5.9	6.3
TOTAL PRIMARY & SECONDARY SLUDGE DISPOSAL COSTS, $/YEAR				
Waste Activated Sludge Thickening	8,600	8,100	7,700	7,100
Anaerobic Digestion	25,200	24,500	23,600	22,500
Vacuum Filtration	104,000	101,000	97,800	92,800
Cake Disposal by Landfill	14,700	14,400	14,000	13,500
Annual Sludge Disposal Costs	152,500	148,000	143,100	135,900
Cents/1000 gallons	7.0	6.8	6.5	6.2
TOTAL TREATMENT SYSTEM ANNUAL COSTS, $/YEAR				
Annual Secondary Costs	113,000	121,600	129,400	137,200
Annual Sludge Disposal Costs	152,500	148,000	143,100	135,900
Annual Common Plant Facilities Costs	78,90	78,900	78,900	78,900
Total Plant Annual Costs	344,400	348,500	351,400	352,000
Cents/1000 gallons	15.7	15.9	16.0	16.1

TABLE 56

ESTIMATED SECONDARY TREATMENT, SLUDGE DISPOSAL, AND TOTAL PLANT COSTS WITH AIR AERATION

30 MGD Treated, 0.8 CF(NTP) of Air/Gallon, or 750 CF(NTP) of Air/Lb. BOD

	Hours Detention Based on Raw Sewage Flow			
	3.0	4.0	5.0	6.0
SECONDARY TREATMENT SYSTEM INVESTMENT,$				
Aeration Tanks	488,000	640,000	788,000	945,000
Sludge Return Pumps & Piping	156,000	156,000	156,000	156,000
Blowers & Diffusers	488,000	488,000	488,000	488,000
Final Sedimentation Tanks	424,000	424,000	424,000	424,000
Total Secondary Investment	1,556,000	1,708,000	1,856,000	2,013,000
Annual Secondary Investment Cost,$/Yr.	116,000	127,000	138,000	150,000
SECONDARY TREATMENT SYSTEM OPERATING & MAINTENANCE COSTS, $/YEAR				
Blower Power	48,000	48,000	48,000	48,000
M & R Costs & Sludge Return Pump Power	76,000	95,000	113,000	131,000
Annual Secondary O & M Costs	124,000	143,000	161,000	179,000
SECONDARY TREATMENT SYSTEM ANNUAL COSTS, $/YEAR				
Secondary Investment	116,000	127,000	138,000	150,000
Secondary O & M	124,000	143,000	161,000	179,000
Annual Secondary Treatment Costs	240,000	270,000	299,000	329,000
Cents/1000 gallons	2.2	2.5	2.7	3.0
TOTAL PRIMARY & SECONDARY SLUDGE DISPOSAL COSTS, $/YEAR				
Waste Activated Sludge Thickening	35,000	33,000	31,000	29,000
Anaerobic Digestion	109,000	105,000	101,000	97,000
Vacuum Filtration	431,000	415,000	401,000	382,000
Incineration	275,000	267,000	259,000	247,000
Annual Sludge Disposal Costs	850,000	820,000	792,000	755,000
Cents/1000 gallons	7.8	7.5	7.2	6.9
TOTAL TREATMENT SYSTEM ANNUAL COSTS, $/YEAR				
Annual Secondary Costs	240,000	270,000	299,000	329,000
Annual Sludge Disposal Costs	850,000	820,000	792,000	755,000
Annual Common Plant Facilities Costs	259,000	259,000	259,000	259,000
Total Plant Annual Costs	1,349,000	1,349,000	1,350,000	1,343,000
Cents/1000 gallons	12.3	12.3	12.3	12.3

TABLE 57

ESTIMATED SECONDARY TREATMENT, SLUDGE DISPOSAL, AND TOTAL PLANT COSTS WITH AIR AERATION

30 MGD Treated, 1.6 CF(NTP) of Air/Gallon, or 1500 CF(NTP) of Air/lb. BOD

	Hours Detention Based on Raw Sewage Flow			
	3.0	4.0	5.0	6.0
SECONDARY TREATMENT SYSTEM INVESTMENT, $				
Aeration tanks	488,000	640,000	788,000	945,000
Sludge Return Pumps & Piping	156,000	156,000	156,000	156,000
Blowers & Diffusers	800,000	800,000	800,000	800,000
Final Sedimentation Tanks	424,000	424,000	424,000	424,000
Total Secondary Investment	1,868,000	2,020,000	2,168,000	2,325,000
Annual Secondary Investment Costs, $/Yr.	139,000	150,000	162,000	173,000
SECONDARY TREATMENT SYSTEM OPERATING & MAINTENANCE COSTS, $/YEAR				
Blower Power	95,000	95,000	95,000	95,000
M & R Costs & Sludge Return Pump Power	84,000	105,000	125,000	145,000
Annual Secondary O & M Costs	179,000	200,000	220,000	240,000
SECONDARY TREATMENT SYSTEM ANNUAL COSTS, $/YEAR				
Secondary Investment	139,000	150,000	162,000	173,000
Secondary O & M	179,000	200,000	220,000	240,000
Annual Secondary Treatment Costs	318,000	350,000	382,000	413,000
Cents/1000 gallons	2.9	3.2	3.5	3.8
TOTAL PRIMARY & SECONDARY SLUDGE DISPOSAL COSTS, $/YEAR				
Waste Activated Sludge Thickening	35,000	33,000	31,000	29,000
Anaerobic Digestion	109,000	105,000	101,000	97,000
Vacuum Filtration	431,000	415,000	401,000	382,000
Incineration	275,000	267,000	259,000	247,000
Annual Sludge Disposal Costs	850,000	820,000	792,000	755,000
Cents/1000 gallons	7.8	7.5	7.2	6.9
TOTAL TREATMENT SYSTEM ANNUAL COSTS, $/YEAR				
Annual Secondary Costs	318,000	350,000	382,000	413,000
Annual Sludge Disposal Costs	850,000	820,000	792,000	755,000
Annual Common Plant Facilities Costs	259,000	259,000	259,000	259,000
Total Plant Annual Costs	1,427,000	1,429,000	1,433,000	1,427,000
Cents/1000 gallons	13.0	13.1	13.1	13.0

TABLE 58

ESTIMATED SECONDARY TREATMENT, SLUDGE DISPOSAL, AND TOTAL PLANT COSTS WITH AIR AERATION

30 MGD Treated, 2.4 CF(NTP) of Air/Gallon, or 2250 CF(NTP) of Air/Lb.BOD

	Hours Detention Based on Raw Sewage Flow			
	3.0	4.0	5.0	6.0
SECONDARY TREATMENT SYSTEM INVESTMENT,$				
Aeration Tanks	488,000	640,000	788,000	945,000
Sludge Return Pumps & Piping	156,000	156,000	156,000	156,000
Blowers & Diffusers	1,050,000	1,050,000	1,050,000	1,050,000
Final Sedimentation tanks	424,000	424,000	424,000	424,000
Total Secondary Investment	2,118,000	2,270,000	2,418,000	2,575,000
Annual Secondary Investment Costs,$/Yr.	158,000	169,000	180,000	192,000
SECONDARY TREATMENT SYSTEM OPERATING & MAINTENANCE COSTS, $/YEAR				
Blower Power	143,000	143,000	143,000	143,000
M & R Costs & Sludge Return Pump Power	92,000	114,000	137,000	159,000
Annual Secondary O & M Costs	235,000	257,000	280,000	302,000
SECONDARY TREATMENT SYSTEM ANNUAL COSTS, $/YEAR				
Secondary Investment	158,000	169,000	180,000	192,000
Secondary O & M	235,000	257,000	280,000	302,000
Annual Secondary Treatment Costs	393,000	426,000	460,000	494,000
Cents/1000 Gallons	3.6	3.9	4.2	4.5
TOTAL PRIMARY & SECONDARY SLUDGE DISPOSAL COSTS, $/YEAR				
Waste Activated Sludge Thickening	35,000	33,000	31,000	29,000
Anaerobic Digestion	109,000	105,000	101,000	97,000
Vacuum Filtration	431,000	415,000	401,000	382,000
Incineration	275,000	267,000	259,000	247,000
Annual Sludge Disposal Costs	850,000	820,000	792,000	755,000
Cents/1000 Gallons	7.8	7.5	7.2	6.9
TOTAL TREATMENT SYSTEM ANNUAL COSTS, $/YEAR				
Annual Secondary Costs	393,000	426,000	460,000	494,000
Annual Sludge Disposal Costs	850,000	820,000	792,000	755,000
Annual Common Plant Facilities Costs	259,000	259,000	259,000	259,000
Total Plant Annual Costs	1,502,000	1,505,000	1,511,000	1,508,000
Cents/1000 Gallons	13.7	13.7	13.8	13.8

TABLE 59

ESTIMATED SECONDARY TREATMENT, SLUDGE DISPOSAL, AND TOTAL PLANT COSTS WITH AIR AERATION

100 MGD Treated, 0.8 CF(NTP) of Air/Gallon, or 750 CF(NTP) of Air/Lb. BOD

	Hours Detention Based on Raw Sewage Flow			
	3.0	4.0	5.0	6.0
SECONDARY TREATMENT SYSTEM INVESTMENT, $				
Aeration Tanks	1,575,000	2,104,000	2,621,000	3,150,000
Sludge Return Pumps & Piping	312,000	312,000	312,000	312,000
Blowers & Diffusers	1,140,000	1,140,000	1,140,000	1,140,000
Final Sedimentation Tanks	1,240,000	1,240,000	1,240,000	1,240,000
Total Secondary Investment	4,267,000	4,796,000	5,313,000	5,842,000
Annual Secondary Investment Costs, $/Yr.	318,000	358,000	396,000	436,000
SECONDARY TREATMENT SYSTEM OPERATING & MAINTENANCE COSTS, $/YEAR				
Blower Power	135,000	135,000	135,000	135,000
M & R Costs & Sludge Return Pump Power	205,000	260,000	313,000	366,000
Annual Secondary O & M Costs	340,000	395,000	448,000	501,000
SECONDARY TREATMENT SYSTEM ANNUAL COSTS, $/YEAR				
Secondary Investment	318,000	358,000	396,000	436,000
Secondary O & M	340,000	395,000	448,000	501,000
Annual Secondary Treatment Costs	658,000	753,000	844,000	937,000
Cents/1000 Gallons	1.8	2.1	2.3	2.6
TOTAL PRIMARY & SECONDARY SLUDGE DISPOSAL COSTS, $/YEAR				
Waste Activated Sludge Thickening	99,000	95,000	89,000	83,000
Anaerobic Digestion	325,000	316,000	304,000	291,000
Vacuum Filtration	1,253,000	1,206,000	1,168,000	1,110,000
Incineration	712,000	691,000	661,000	642,000
Annual Sludge Disposal Costs	2,389,000	2,308,000	2,222,000	2,126,000
Cents/1000 Gallons	6.5	6.3	6.1	5.8
TOTAL TREATMENT SYSTEM ANNUAL COSTS, $/YEAR				
Annual Secondary Costs	658,000	753,000	844,000	937,000
Annual Sludge Disposal Costs	2,389,000	2,308,000	2,222,000	2,126,000
Annual Common Plant Facilities Costs	705,000	705,000	705,000	705,000
Total Plant Annual Costs	3,752,000	3,766,000	3,771,000	3,768,000
Cents/1000 Gallons	10.3	10.3	10.3	10.3

TABLE 60

ESTIMATED SECONDARY TREATMENT, SLUDGE DISPOSAL, AND TOTAL PLANT COSTS WITH AIR AERATION

100 MGD Treated, 1.6 CF(NTP) of Air/Gallon, or 1500 CF(NTP) of Air/lb. BOD

	Hours Detention Based on Raw Sewage Flow			
	3.0	4.0	5.0	6.0
SECONDARY TREATMENT SYSTEM INVESTMENT,$				
Aeration Tanks	1,575,000	2,104,000	2,621,000	3,150,000
Sludge Return Pumps & Piping	312,000	312,000	312,000	312,000
Blowers & Diffusers	1,910,000	1,910,000	1,910,000	1,910,000
Final Sedimentation Tanks	1,240,000	1,240,000	1,240,000	1,240,000
Total Secondary Investment	5,037,000	5,566,000	6,083,000	6,612,000
Annual Secondary Investment Cost,$/Yr.	376,000	415,000	453,000	493,000
SECONDARY TREATMENT SYSTEM OPERATING & MAINTENANCE COSTS, $/YEAR				
Blower Power	269,000	269,000	269,000	269,000
M & R Costs & Sludge Return Pump Power	227,000	288,000	347,000	405,000
Annual Secondary O & M Costs	496,000	557,000	616,000	674,000
SECONDARY TREATMENT SYSTEM ANNUAL COSTS, $/YEAR				
Secondary Investment	376,000	415,000	453,000	493,000
Secondary O & M	496,000	557,000	616,000	674,000
Annual Secondary Treatment Costs	872,000	972,000	1,069,000	1,167,000
Cents/1000 Gallons	2.4	2.7	2.9	3.2
TOTAL PRIMARY & SECONDARY SLUDGE DISPOSAL COSTS $/YEAR				
Waste Activated Sludge Thickening	99,000	95,000	89,000	83,000
Anaerobic Digestion	325,000	316,000	304,000	291,000
Vacuum Filtration	1,253,000	1,206,000	1,168,000	1,110,000
Incineration	712,000	691,000	661,000	642,000
Annual Sludge Disposal Costs	2,389,000	2,308,000	2,222,000	2,126,000
Cents/1000 Gallons	6.5	6.3	6.1	5.8
TOTAL TREATMENT SYSTEM ANNUAL COSTS, $/YEAR				
Annual Secondary Costs	872,000	972,000	1,069,000	1,167,000
Annual Sludge Disposal Costs	2,389,000	2,308,000	2,222,000	2,126,000
Annual Common Plant Facilities Costs	705,000	705,000	705,000	705,000
Total Plant Annual Costs	3,966,000	3,985,000	3,996,000	3,998,000
Cents/1000 Gallons	10.9	10.9	10.9	11.0

TABLE 61

ESTIMATED SECONDARY TREATMENT, SLUDGE DISPOSAL, AND TOTAL PLANT COSTS WITH AIR AERATION

100 MGD Treated, 2.4 CF(NTP) of Air/Gallon, or 2250 CF(NTP) of Air/Lb.BOD

	Hours Detention Based on Raw Sewage Flow			
	3.0	4.0	5.0	6.0
SECONDARY TREATMENT SYSTEM INVESTMENT, $				
Aeration Tanks	1,575,000	2,104,000	2,621,000	3,150,000
Sludge Retun Pumps & Piping	312,000	312,000	312,000	312,000
Blowers & Diffusers	2,710,000	2,710,000	2,710,000	2,710,000
Final Sedimentation Tanks	1,240,000	1,240,000	1,240,000	1,240,000
Total Secondary Investment	5,837,000	6,366,000	6,883,000	7,412,000
Annual Secondary Investment Cost, $/Yr.	435,000	475,000	513,000	553,000
SECONDARY TREATMENT SYSTEM OPERATING & MAINTENANCE COSTS, $/YEAR				
Blower Power	404,000	404,000	404,000	404,000
M & R Costs & Sludge Return Pump Power	248,000	315,000	380,000	445,000
Annual Secondary O & M Costs	652,000	719,000	784,000	849,000
SECONDARY TREATMENT SYSTEM ANNUAL COSTS, $/YEAR				
Secondary Investment	435,000	475,000	513,000	553,000
Secondary O & M	652,000	719,000	784,000	849,000
Annual Secondary Treatment Cost	1,087,000	1,194,000	1,297,000	1,402,000
Cents/1000 Gallons	3.0	3.3	3.6	3.8
TOTAL PRIMARY & SECONDARY SLUDGE DISPOSAL COSTS, $/YEAR				
Waste Activated Sludge Thickening	99,000	95,000	89,000	83,000
Anaerobic Digestion	325,000	316,000	304,000	291,000
Vacuum Filtration	1,253,000	1,206,000	1,168,000	1,110,000
Incineration	712,000	691,000	661,000	642,000
Annual Sludge Disposal Costs	2,389,000	2,308,000	2,222,000	2,126,000
Cents/1000 Gallons	6.5	6.3	6.1	5.8
TOTAL TREATMENT SYSTEM ANNUAL COSTS, $/YEAR				
Annual Secondary Costs	1,087,000	1,194,000	1,297,000	1,402,000
Annual Sludge Disposal Costs	2,389,000	2,308,000	2,222,000	2,126,000
Annual Common Plant Facilities Costs	705,000	705,000	705,000	705,000
Total Plant Annual Costs	4,181,000	4,207,000	4,224,000	4,233,000
Cents/1000 Gallons	11.5	11.5	11.6	11.6

Figures 26 through 29 compare total treatment costs for the oxygenation and aeration systems for each of the four plant sizes as a function of nominal aeration detention time (raw wastewater flow only) for the three assumed aeration rates. It is apparent from inspection of these figures that the oxygenation process offers significant cost reductions over the range of plant sizes from 6 MGD to 100 MGD when compared with the air aeration systems.

When compared with the lowest air aeration rate of 0.8 cf(NTP)/gal. of wastewater treated, the estimated savings in total treatment costs with oxygenation range from 11-15% at the 6 MGD size to 17-25% at the 100 MGD size. Compared with an air aeration rate of 1.6 cf(NTP)/gal. of wastewater treated, the oxygenation system offers a savings in total treatment cost of 17-21% at the 6 MGD size to 22-29% at the 100 MGD size. At the 1 MGD plant size, the total treatment costs are about equal at an air aeration rate of 1.6 cf(NTP)/gal. of wastewater treated. For the 1 MGD plant size, the oxygenation system shows 8-10% higher total treatment costs than an air system with an aeration rate of 0.8 cf(NTP)/gal. of wastewater treated and about 8-10% lower treatment costs at an air aeration rate of 2.4 cf(NTP)/gal. of wastewater treated.

Another meaningful comparison of the relative cost effectiveness of the two processes can be gained by consideration of the total treatment system costs, exclusive of the common plant facilities cost. These common plant facilities costs would be incurred for most types of secondary treatment systems and are unaffected by choice of aeration process. Table 62 compares the combined secondary treatment and sludge disposal costs only for the oxygenation system and the air system at an air aeration rate of 1.6 cf(NTP)/gal. of wastewater treated.

As shown in Table 62, the oxygenation system cost estimates indicate no savings in treatment costs at the 1 MGD plant size. At the 6, 30, and 100 MGD plant sizes, cost reductions of 23-27%, 26-33%, and 27-32%, respectively, are indicated for the oxygenation system. These cost differentials

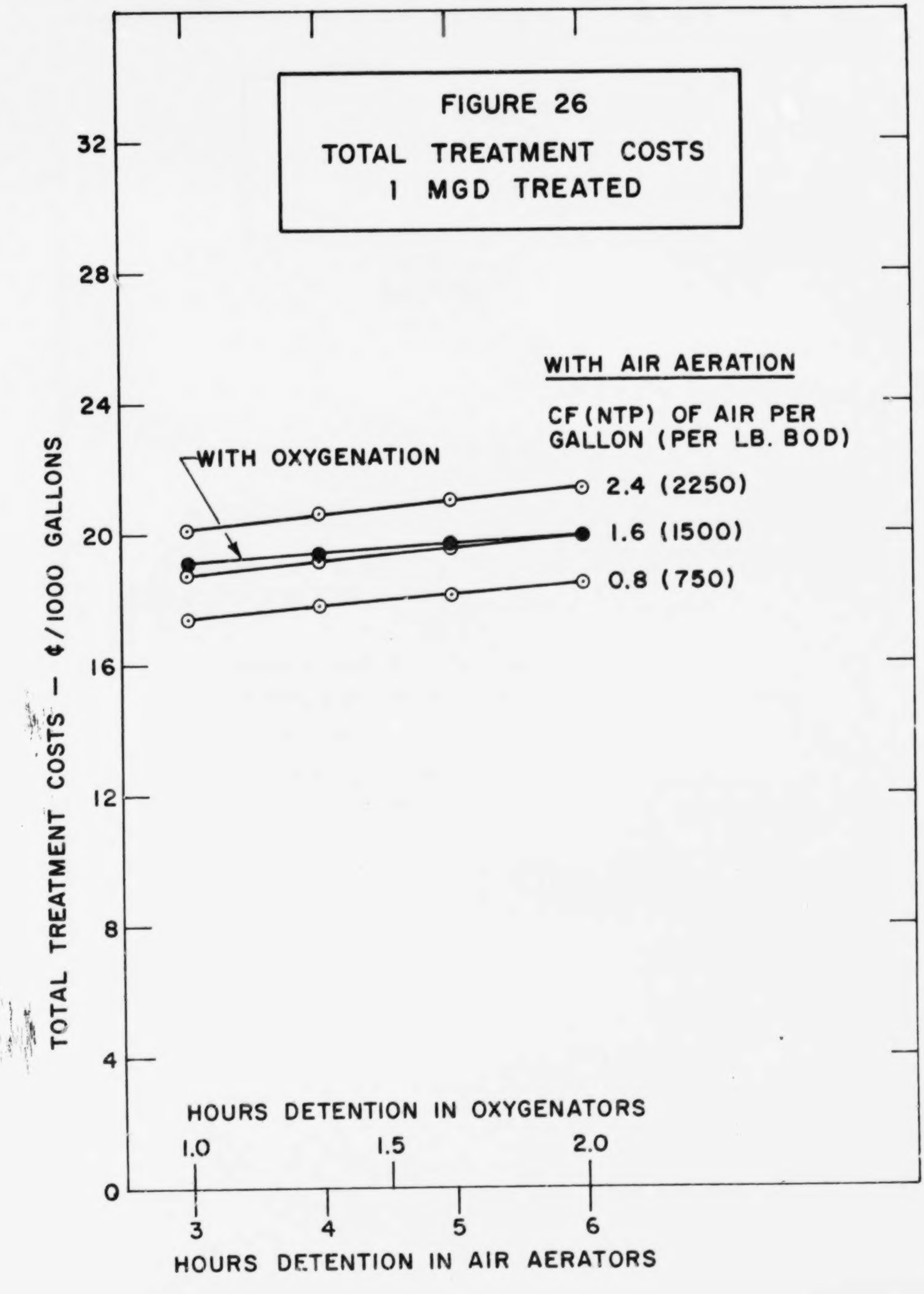


FIGURE 26
TOTAL TREATMENT COSTS
1 MGD TREATED

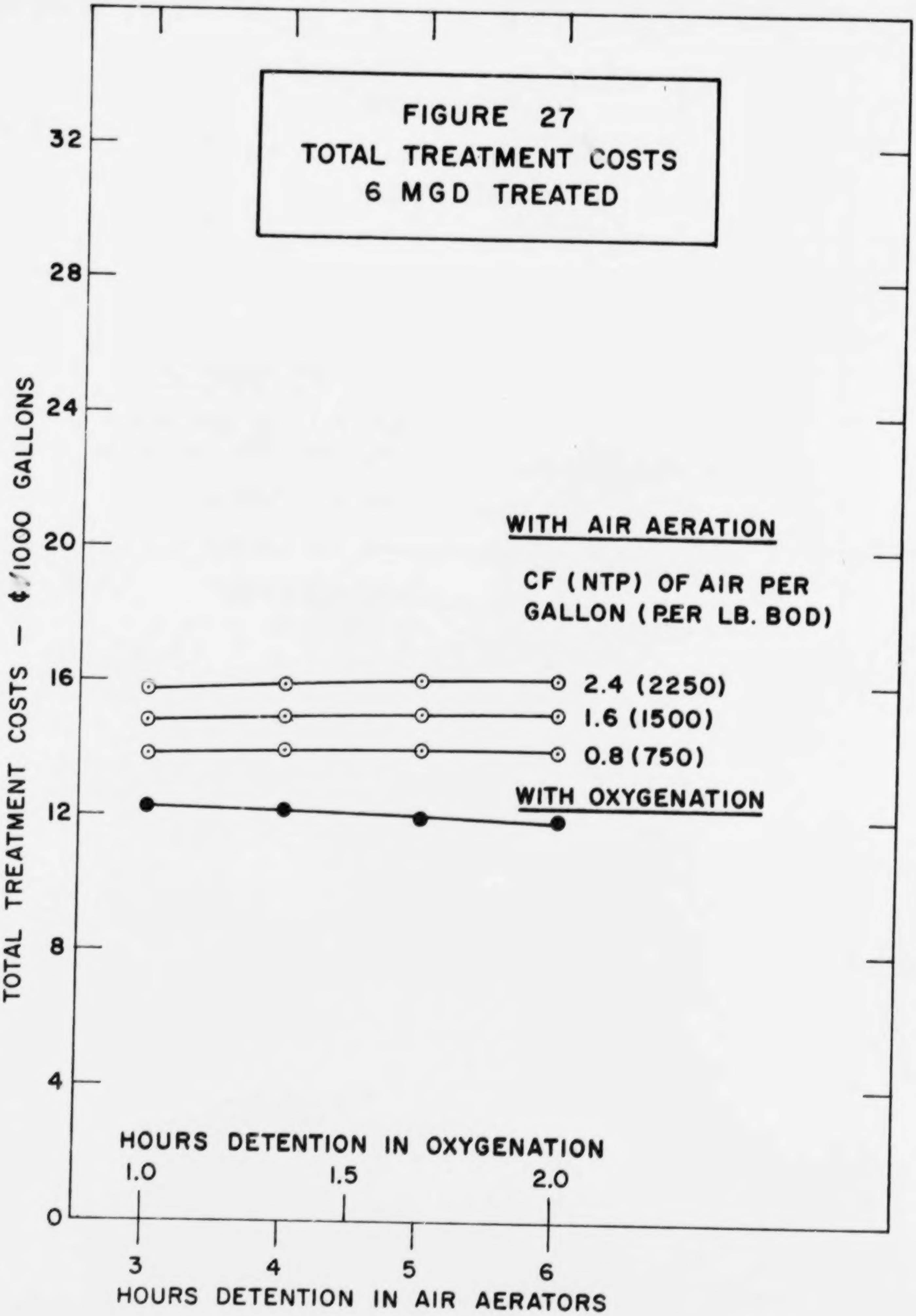


FIGURE 27
TOTAL TREATMENT COSTS
6 MGD TREATED

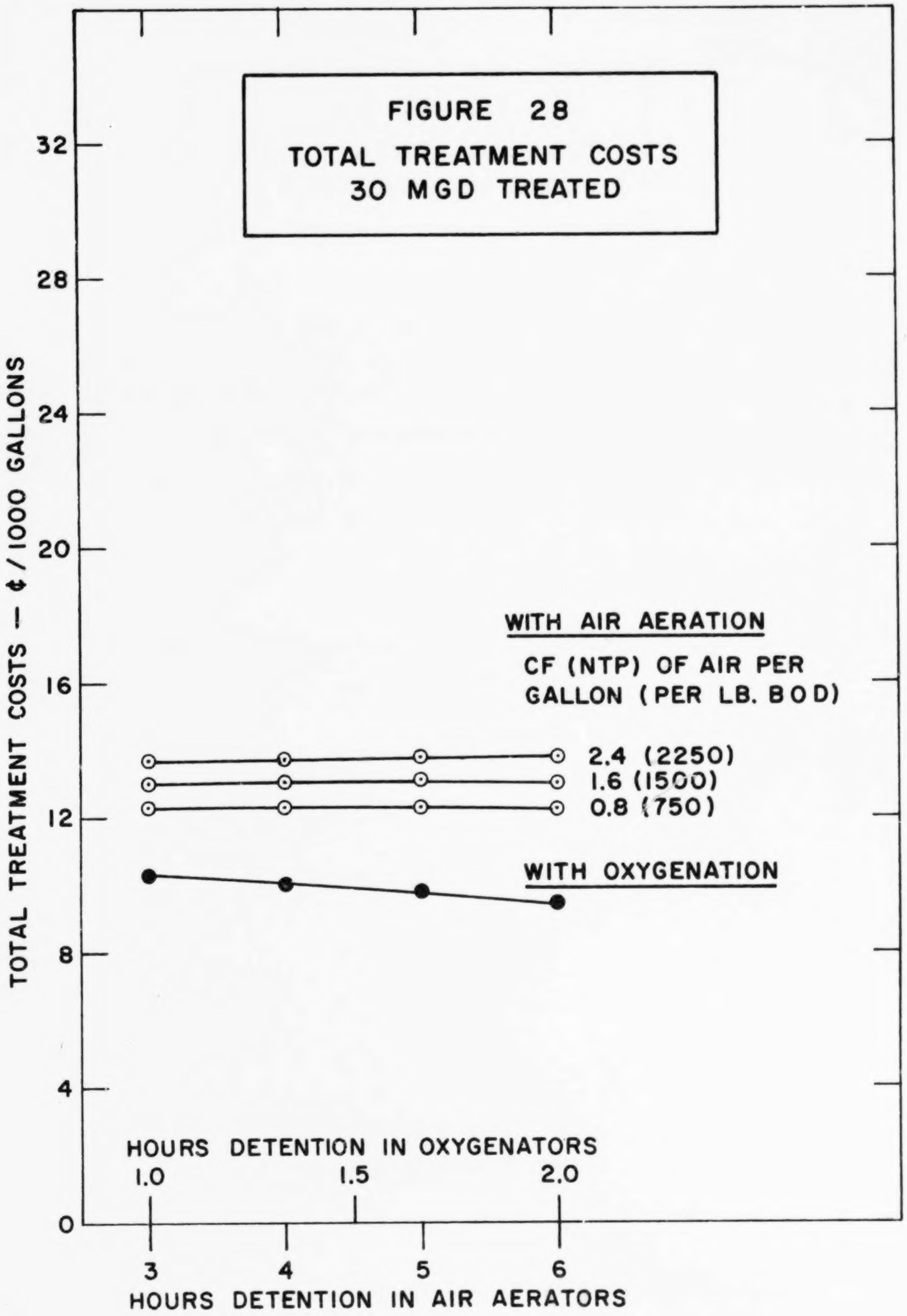


FIGURE 28
TOTAL TREATMENT COSTS
30 MGD TREATED

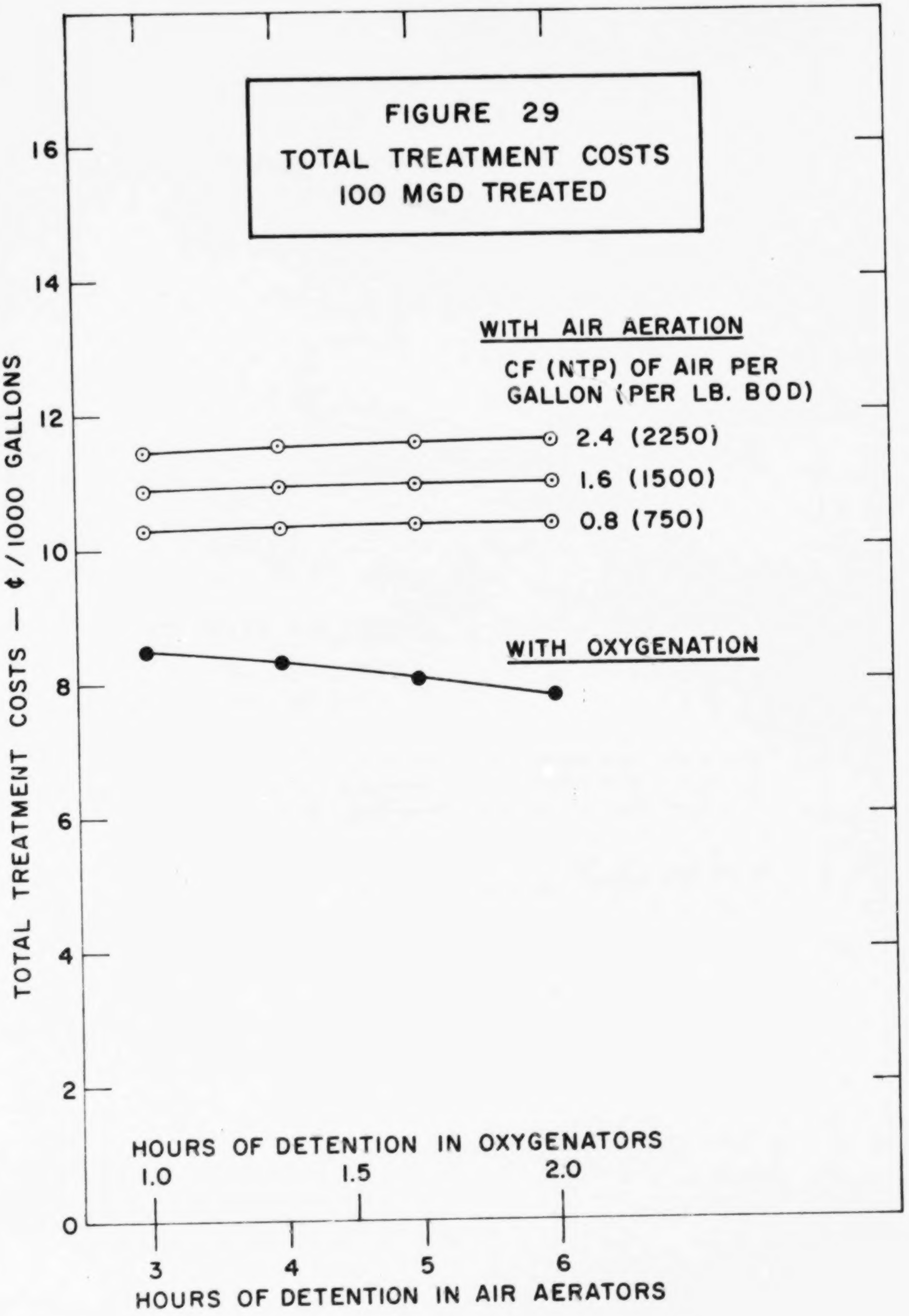


FIGURE 29
TOTAL TREATMENT COSTS
100 MGD TREATED

TABLE 62

COMPARISON OF COMBINED SECONDARY TREATMENT AND SLUDGE DISPOSAL COSTS AS CENTS PER THOUSAND GALLONS OF WASTEWATER TREATED

AIR AERATION

1.6 cf(NTP)/gal. of Wastewater Treated

Plant Size (MGD)	Detention Time (Hrs.)			
	3.00	4.00	5.00	6.00
1	11.5	11.9	12.3	12.6
6	11.2	11.3	10.5	11.5
30	10.7	10.7	10.7	10.7
100	8.9	9.0	9.0	9.0

OXYGENATION

Plant Size (MGD)	Detention Time (Hrs.)			
	1.00	1.33	1.66	2.00
1	11.9	12.1	12.4	12.7
6	8.6	8.6	8.4	8.3
30	7.9	7.7	7.5	7.1
100	6.6	6.4	6.1	5.8

for 6, 30, and 100 MGD plant sizes are the result of reductions in both the secondary treatment system costs and the sludge disposal costs for the oxygenation system. Of these cost differentials, about one-third is attributable to cost reductions in the secondary treatment system and about two-thirds to cost reductions in sludge disposal. The sludge disposal cost reductions with oxygenation are in turn primarily a result of the reduced quantity of waste activated sludge produced, as illustrated in the work presented herein. Additional work is, however, planned to further substantiate the quantity of waste activated sludge produced by the oxygenation system under various loading conditions with and without primary treatment. The data used in support of the economic comparisons here is the most representative data available and is especially meaningful considering that both the air and the oxygenation systems were evaluated treating the same influent wastewater.

An additional relative cost factor, not included in the comparison of Tables 46 through 61, is the potential land cost savings attainable with the oxygenation system. The substantial reduction in the aeration tankage requirements realized with the oxygenation system may result in a 40 ± 10% reduction in the land area required for the secondary treatment system. In many metropolitan treatment facilities, this reduction in land requirements can appreciably affect the entire secondary treatment system investment costs. The reduced land requirements can be particularly important in increasing the capacity and/or treatment effectiveness of an existing air aerated activated sludge treatment system by conversion to an oxygenation system. In many cases, the plant capacity can be increased 2-3 fold with no additional aeration tankage required.

Several other features of the oxygenation system are of considerable importance and are also not reflected in these economic comparisons. These include the following: 1) Improved control of mixed-liquor dissolved oxygen concentrations and enhanced capability for responding to wide variations in oxygen demand, 2) High dissolved oxygen concentrations in the mixed-liquor and secondary effluent, 3) Elimination of odor problems because of the small quantity of exhaust gas and the covered aeration tank design.

The improved control of mixed-liquor dissolved oxygen concentrations also results in more uniform and consistently high quality effluents. A simple pressure control system automatically determines the required oxygen gas feed rate and moderates the oxygen gas production rate to meet the oxygen demands of the wastewater being treated. Therefore, the oxygen production power cost, which represents a substantial fraction of the oxygenation system total power costs, can be optimized at all times. This is accomplished without the need for direct mixed-liquor dissolved oxygen measurements, total organic carbon analysis or other suitable measurements to determine the incoming waste strength.

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TABLE I

FEED WASTE, MIXED-LIQUOR SUPERNATANT AND CLARIFIER EFFLUENT SPECIAL ANALYSIS

PHASE I OPERATION

		Week of Operation											
		5/12/69		5/19/69		5/26/69		6/2/69		6/9/69		6/16/69	
Parameter	Source	Air	Oxygen	Air	Oxygen	Air	Oxygen	Air	Oxygen	Air	Oxygen	Air	Oxygen
COD	Feed Waste												
mg/l	Inlet	261.0	216.0	57.0	32.0	351.0	387.0	122.0	84.0	214.0	58.0	94.0	56.0
	Middle		267.0		208.0		75.0		61.0		32.0		175.0
	Outlet	246.0	183.0	41.0	32.0	112.0	96.0	77.0	59.0	89.0	52.0	99.0	72.0
	Clar.Eff.												
BOD	Feed Waste	101.0	101.0	98.0	98.0	445.0	445.0	187.0	187.0	153.0	153.0	157.0	157.0
mg/l	Inlet	35.0	27.0	40.0	27.0	180.0	240.0	50.0	32.0	80.0	38.0	53.0	26.0
	Middle		29.0		12.0		15.0		15.0		24.0		12.0
	Outlet	19.0	11.0	15.0	12.0	23.0	11.0	19.0	21.0	60.0	29.0	31.0	15.0
	Clar.Eff.	9.0	7.0	9.0	5.0	6.0	6.0	12.0	11.0	22.0	13.0	16.0	4.0
Total	Feed Waste												
Phosphate	Inlet	5.0	5.0	4.0	4.0	8.0	9.0	4.0	3.0	5.0	8.0	6.0	7.0
mg/l	Middle		8.0		3.0		6.0		3.0		5.0		6.0
	Outlet	6.0	6.0	3.0	3.0	5.0	6.0	3.0	3.0	7.0	7.0	8.0	8.0
	Clar.Eff.												
NH_3-N	Feed Waste												
mg/l	Inlet	23.0	26.0	16.0	18.0	34.0	40.0	6.0	5.0	27.0	35.0	28.0	32.0
	Middle		23.0		8.0		27.0		4.0		15.0		18.0
	Outlet	17.0	6.0	8.0	3.0	27.0	6.0	4.0	3.0	15.0	7.0	19.0	8.0
	Clar.Eff.												
NO_2-N	Feed Waste												
mg/l	Inlet	0.6	0.2	0.8	0.9		0.0	0.0	0.5	0.0	0.1	0.0	0.1
	Middle		0.4		0.4		0.9		0.7		0.9		0.9
	Outlet	0.8	0.3	0.5	0.5	0.0	1.2	0.0	0.8	0.0	0.5	0.0	1.1
	Clar.Eff.												
NO_3-N	Feed Waste												
mg/l	Inlet	0.0	0.9	0.5	0.9	0.1	0.1	0.1	1.2	0.0	0.1	0.0	0.0
	Middle		0.7		1.2		1.0		2.0		0.5		1.6
	Outlet	0.3	0.9	1.1	1.0	0.1	1.4	0.0	1.9	0.0	0.4	0.0	1.5
	Clar.Eff.												
TKN	Feed Waste												
mg/l	Inlet	33.0	45.0	23.0	24.0	60.0	70.0	40.0	45.0	30.0	39.0	34.0	37.0
	Middle		29.0		14.0		33.0		32.0		19.0		21.0
	Outlet	22.0	25.0	15.0	16.0	33.0	32.0	28.0	25.0	18.0	14.0	25.0	24.0
	Clar.Eff.												

TABLE II

FEED WASTE, MIXED-LIQUOR SUPERNATANT AND CLARIFIER EFFLUENT SPECIAL ANALYSIS

PHASE II OPERATION

Parameter	Source	Week of Operation 7/28/69 Oxygen	8/4/69 Oxygen	8/11/69 Oxygen	8/18/69 Oxygen	8/25/69 Oxygen
COD mg/1	Feed Waste					
	Inlet	154.0	136.0	189.0	81.0	93.0
	Middle	171.0	104.0	203.0	163.0	137.0
	Outlet	146.0	123.0	158.0	133.0	159.0
	Clar.Eff.					
BOD mg/1	Feed Waste	165.0	285.0	201.0	195.0	284.0
	Inlet	90.0	55.0	55.0	50.0	63.0
	Middle	62.0	43.0	27.0	72.0	43.0
	Outlet	41.0	45.0	36.0	37.0	32.0
	Clar.Eff.	29.0	21.0	7.0		29.0
Total Phosphate mg/1	Feed Waste					
	Inlet		10.0	11.0	9.0	12.0
	Middle		9.0	9.0	8.0	11.0
	Outlet		9.0	9.0	8.0	11.0
	Clar.Eff.					
NH_3-N mg/1	Feed Waste					
	Inlet		22.0	24.0	20.0	23.0
	Middle		22.0	20.0	20.0	24.0
	Outlet		23.0	19.0	21.0	24.0
	Clar.Eff.					
NO_2-N mg/1	Feed Waste					
	Inlet		0.0	0.0	0.0	
	Middle		0.0	0.0	0.0	
	Outlet		0.0	0.0	0.0	
	Clar.Eff.					
NO_3-N mg/1	Feed Waste					
	Inlet		0.1	0.0	0.1	0.1
	Middle		0.1	0.1	0.1	0.1
	Outlet		0.1	0.1	0.1	0.1
	Clar.Eff.					
TKN mg/1	Feed Waste					
	Inlet		30.0	34.0	29.0	38.0
	Middle		32.0	30.0	26.0	36.0
	Outlet		31.0	28.0	26.0	35.0
	Clar.Eff.					

TABLE III

FEED WASTE, MIXED-LIQUOR SUPERNATANT AND CLARIFIER EFFLUENT SPECIAL ANALYSIS

PHASE III OPERATION

		Week of Operation													
		10/20/69		10/27/69		11/3/69		9/8/69		9/15/69		9/22/69		9/29/69	
Parameter	Source	Air	Oxygen	Air	Oxygen	Air	Oxygen	Air	Oxygen	Air	Oxygen	Air	Oxygen	Air	Oxygen
COD mg/l	Feed Waste														
	Inlet	150.0	123.0	123.0	87.0	184.0	87.0	213.0		281.0	188.0	181.0	118.0	219.0	142.0
	Middle	149.9	88.0	151.0	172.0	248.0	78.0	179.0	139.0	190.0	123.0	212.0	172.0	146.0	187.0
	Outlet	129.0	83.0	105.0	193.0	182.0	112.0	163.0	151.0	140.0	120.0	134.0	162.0	154.0	185.0
	Clar.Eff.														
BOD mg/l	Feed Waste	294.0	294.0	233.0	233.0	155.0	155.0			246.0	246.0	229.0	229.0	288.0	288.0
	Inlet	91.0	34.0	45.0	41.0	150.0	66.0	111.0	67.0	97.0	43.0	168.0	57.0	173.0	121.0
	Middle	38.0	24.0	43.0	23.0	105.0	50.0	73.0	38.0	53.0	29.0	124.0	43.0	90.0	52.0
	Outlet	32.0	19.0	36.0	27.0	74.0	46.0	46.0	40.0	46.0	19.0		42.0		41.0
	Clar.Eff.	29.0	23.0	24.0	12.0	29.0	18.0			28.0	17.0	22.0	13.0	38.0	20.0
Phosphate mg/l	Feed Waste														
	Inlet	11.0	14.0	15.0	16.0	8.0	9.0	10.9	13.0	8.0	12.0	13.0	15.0	10.0	14.0
	Middle	11.0	14.0	15.0	16.0	8.0	9.0	9.0	12.0	8.0	13.0	13.0	14.0	10.0	14.0
	Outlet	10.0	14.0	15.0	16.0	8.0	8.0	9.0	13.0	8.0		13.0	12.0	10.0	16.0
	Clar.Eff.														
NH_3-N mg/l	Feed Waste														
	Inlet	34.0	37.0	42.0	44.0	32.0	30.0	31.0	37.0	33.0	36.0	33.0	27.0	35.0	37.0
	Middle	33.0	38.0	40.0	44.0	32.0	29.0	31.0	38.0	32.0	35.0	32.0	31.0	34.0	35.0
	Outlet	32.0	37.0	37.0	42.0	30.0	28.0	30.0	39.0	31.0	33.0	29.0	30.0	31.0	32.0
	Clar.Eff.														
NO_2-N mg/l	Feed Waste														
	Inlet	0.1	0.0	0.1	0.6	0.0	1.0	0.0	0.0	0.0	0.0	0.0	0.0	0.1	0.1
	Middle	0.0	0.0	0.0	0.9	0.0	1.3	0.0	0.1	0.0	0.1	0.0	0.0	0.0	0.3
	Outlet	0.0	1.0	0.0	1.1	0.0	1.6	0.0	0.1	0.0	0.1	0.0	0.0	0.0	0.6
	Clar.Eff.														
NO_3-N mg/l	Feed Waste														
	Inlet	0.1	0.1	0.1	1.2	0.1	0.4	0.1	0.1	0.1	0.2	0.1	0.0	0.1	0.1
	Middle	0.1	0.1	0.1	1.3	0.1	0.8	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1
	Outlet	0.1	-0.6	0.1	1.4	0.1	1.1	0.1	0.2	0.1	0.1	0.1	0.0	0.1	0.1
	Clar.Eff.														
TKN mg/l	Feed Waste														
	Inlet	41.0	44.0	54.0	55.0	41.0	38.0	39.0	42.0	43.0	45.0	48.0	39.0	48.0	55.0
	Middle	42.0	45.0	48.0	54.0	37.0	36.0	38.0	41.0	42.0	42.0	49.0	47.0	48.0	47.0
	Outlet	39.0	45.0	47.0	52.0	37.0	34.0	38.0	43.0	41.0	40.0	44.0	42.0	49.0	46.0
	Clar.Eff.														

END